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Studies of Flue Gas Desulfurization at Louisville Gas and Electric´s Paddy´s Run Station: Volume II. Review and Analysis of Commercial Operations

Radian Corp.
Austin, TX

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STUDIES OF FLUE GAS DESULFURIZATION AT LOUISVILLE GAS & ELECTRIC'S PADDY'S RUN STATION:

VOLUME II.
REVIEW AND ANALYSIS OF COMMERCIAL OPERATIONS

Prepared for:
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16. ABSTRACT The report gives results of tests to determine the technical factors accounting for the success of the flue gas desulfurization (FGD) system at Louisville Gas and Electric Co.'s Paddy's Run Unit 6. (Between its start-up in the Spring of 1973 and the Fall of 1976, the Unit 6 FGD system logged almost 4000 hours of operation without any major process or mechanical problem.) Test results indicate that carbide lime (a waste by-product of acetylene manufacture using calcium carbide), used at Unit 6, contains oxidation inhibitors that prevent gypsum scaling in the FGD scrubber. Commercial lime testing without these inhibitors resulted in gypsum scaling. However, adding magnesium (3000 ppm in the slurry liquor) reduced sulfite dissolution in the scrubber, thereby reducing sulfite oxidation and allowing scale-free operation. Increasing the chloride level increased sulfite oxidation; an increase in magnesium addition offset the effects of higher chloride levels. Lime systems can be operated successfully with low reaction tank residence times because of the rapid dissolution of lime. The size of the reaction tank and the lime addition point affect the crystal structure of the scrubber waste solids.

17. KEY WORDS AND DOCUMENT ANALYSIS

a. DESCRIPTORS		b. IDENTIFIERS/OPEN ENDED TERMS	c. COSATI Field/Group	
Pollution	Gypsum	Pollution Control	13B	08G
Flue Gases	Scaling	Stationary Sources	21B	11C
Desulfurization	Magnesium	Lime	07A, 07D	
Carbides	Sulfites		07B	
Calcium Oxides	Chlorides			
Scrubbers	Wastes		13I	14G
	Crystal Structure			20B

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1.0 INTRODUCTION

EPA plans to conduct a program with Louisville Gas and Electric Company (LG&E) to characterize the operation of LG&E's Paddy's Run No. 6 SO_2 control system. The Paddy's Run No. 6 unit is a full-scale (65 Mw) lime scrubbing system and is one of the first successful commercial scrubber systems in operation.

The goal of MOSC Task No. 30 was to review and analyze previous operating data so that a firm basis for test program development could be established. The following specific objectives of the task were outlined:

1) describe the LG&E process and process flow sheet

2) summarize the system's operating history to date,

3) assess the completeness and quality of available process data, and

4) conduct an engineering analysis of the Paddy's Run SO_2 control system based on existing data

Section 2.0 provides a general process description for a lime scrubbing system with emphasis placed on those situations that are directly applicable to the LG&E scrubbing system. Also included in Section 2.0, is a functional description of the system. This involves an equipment description, a discussion of operating and control strategies, and a listing of the important process flow rates.

A summary of the operating history follows in Section 3.0. General history is furnished as well as specific unit availabilities and SO_2 removal efficiencies. As a general rule, the system was 90% available and exhibited from 80 to 90% SO_2 removal. Process and equipment difficulties which were encountered during operation are also addressed.

Section 4.0 provides a detailed assessment of data gathered during the week November 23-30, 1973. An engineering analysis of the data for this period is presented which shows that the oxidation rate for the system averaged about five percent and that the system operated in a mode which was subsaturated with respect to calcium sulfate. Also, it is shown that significant quantities of calcium sulfite were apparently dissolved in the LG&E scrubber.

Finally, the results of a computer process simulation of the LG&E system are presented in Section 5.0. It is noted that the LG&E system cannot be accurately modeled without the kinetics of co-precipitation being better defined. However, important insight into the operation and process chemistry of this system was derived from the simulation. The simulation results emphasized the importance of calcium sulfite dissolution in the scrubber and low system oxidation rates. Sulfite dissolution in the scrubber is necessary to obtain the high levels of SO_2 removal.

It is concluded that the factors which are responsible for the low oxidation levels and the co-precipitation phenomenon which are observed in the LG&E system need to be investigated further since these factors affect significantly the calcium sulfate scaling potential of lime-limestone SO_2 scrubbing systems.

2.0 PROCESS AND SYSTEM DESCRIPTION

The process and system description for Louisville Gas and Electric's Paddy's Run No. 6 SO_2 scrubbing system is divided into two subsections. The first describes the general process chemistry of a lime scrubbing system. The second subsection gives a functional description of the LG & E system including operating and control strategies, description of equipment and materials of construction, and important flow parameters.

2.1 General Process Chemistry

The general process chemistry of a lime scrubbing system is discussed in this subsection. The important chemical species and reactions and their effects on SO_2 removal are presented. Difficulties generally encountered in this type of system are discussed, and the specific operating conditions at LG&E pertinent to these problems are outlined.

2.1.1 Overall Reactions

The chemical objective of a lime wet scrubbing system is to promote the overall reaction of SO_2 with lime to yield solid calcium sulfite as the product. The overall reaction is shown in equation 2-1.

$$SO_2(g) + CaO(s) + \tfrac{1}{2}H_2O \rightarrow CaSO_3 \cdot \tfrac{1}{2}H_2O \qquad (2\text{-}1)$$

The calcium sulfite is only slightly soluble in the scrubbing liquor so precipitation and disposal are possible.

Some carbon dioxide is absorbed from the flue gas. The CO_2 reacts in a similar manner, forming calcium carbonate.

$$CO_2(g) + CaO(s) \rightarrow CaCO_3 \qquad (2\text{-}2)$$

In addition to the SO_2 and CO_2, some oxygen is absorbed which leads to oxidation of the absorbed SO_2. The reaction for this step is:

$$SO_2(g) + \tfrac{1}{2}O_2(g) + CaO(s) + 2H_2O \rightarrow CaSO_4 \cdot 2H_2O(s) \qquad (2\text{-}3)$$

The level of oxidation is an important parameter since it can affect the scaling tendency of the system as well as the settling and dewatering characteristics of the waste solids. The importance of this parameter is discussed further in Section 2.1.3.

2.1.2 Intermediate Reaction Steps

The overall process reactions discussed above involve a number of intermediate steps, some of which are rate limiting. The individual reaction steps are discussed below.

The first step in the process is vapor-liquid mass transfer, which must occur for the overall reaction to proceed. The gaseous species - SO_2, CO_2, and O_2 - must be absorbed by the scrubbing liquor.

$$SO_2(g) \rightarrow SO_2(aq) \qquad (2\text{-}4)$$
$$CO_2(g) \rightarrow CO_2(aq) \qquad (2\text{-}5)$$
$$O_2(g) \rightarrow O_2(aq) \qquad (2\text{-}6)$$

Water is simultaneously transferred to the gas.

$$H_2O(aq) \rightarrow H_2O(g) \qquad (2\text{-}7)$$

The vapor-liquid mass transfer is encouraged by (1) a large gas-liquid interfacial area (2) gas and liquid side agitation, (3) a large residence time of gas-liquid contact, and (4) a sufficient driving force. The first three of these are dependent on the scrubber design. The last item is governed by the process chemistry.

The driving force for sorption of acid species from the gas is maintained by the presence of basic species in the liquid phase. SO_2 can react with a number of different basic species as shown in the following equations.

$$SO_2(aq) + OH^- \rightarrow HSO_3^- \qquad (2\text{-}8)$$

$$SO_2(aq) + CO_3^= \rightarrow SO_3^= + CO_2(aq) \qquad (2\text{-}9)$$

$$SO_2(aq) + HCO_3^- \rightarrow HSO_3^- + CO_2(aq) \qquad (2\text{-}10)$$

$$SO_2(aq) + H_2O + SO_3^= \rightarrow 2HSO_3^- \qquad (2\text{-}11)$$

The basic species which participate in the above reactions (e.g., OH^-, HCO_3^-, $SO_3^=$, or $CO_3^=$) can either be in the scrubber feed liquor or can result from the dissolution of solids. The dissolution of solid calcium hydroxide increases the hydroxyl ion concentration, the dissolution of solid calcium carbonate increases the carbonate ion concentration, and the dissolution of calcium sulfite increases the sulfite ion concentration.

$$Ca(OH)_2(s) \rightarrow Ca^{++} + 2OH^- \qquad (2\text{-}12)$$

$$CaCO_3(s) \rightarrow Ca^{++} + CO_3^= \qquad (2\text{-}13)$$

$$CaSO_3 \cdot \tfrac{1}{2}H_2O(s) \rightarrow Ca^{++} + SO_3^= + \tfrac{1}{2}H_2O \qquad (2\text{-}14)$$

Although under normal conditions lime dissolution would proceed essentially to completion in the hold tank, each of the above reactions is desirable from the standpoint of improving SO_2 removal efficiency, and each can occur to some extent in the scrubber. Lime and carbonate dissolutions in the scrubber can cause scaling problems. (This will be discussed further in Subsection 2.1.3) For this reason, the majority of the basic species which participate in reactions 2-8 to 2-11 above must be provided as either dissolved species in the clear liquor or by calcium sulfite solids dissolution.

Solid-liquid mass transfer (dissolution and precipitation) reactions are enhanced by (1) large solid-liquid interfacial area, (2) reasonable liquid phase agitation, (3) long solid liquid contact residence time, and (4) sufficient driving force. The first three items are determined by the equipment design. The driving force for solid-liquid mass transfer is determined by the process chemistry. Driving forces for the dissolution reactions 2-12, 2-13, and 2-14 are greatest when the liquor is most acidic (i.e., exiting the scrubber). Dissolution driving forces are less in the hold tank where the liquor is more basic.

Precipitation reactions are a second type of important solid-liquid reactions which take place in lime scrubbing systems. The sulfite and sulfate precipitation steps are shown in equations 2-15 and 2-16. Calcium sulfate co-precipitation with calcium sulfite is represented by equation 2-17.

$$Ca^{++} + SO_3^{=} + \tfrac{1}{2}H_2O \rightarrow CaSO_3 \cdot \tfrac{1}{2}H_2O(s) \qquad (2\text{-}15)$$

$$Ca^{++} + SO_4^{=} + 2H_2O \rightarrow CaSO_4 \cdot 2H_2O(s) \qquad (2\text{-}16)$$

$$(1 + x)Ca^{++} + SO_3^{=} + xSO_4^{=} + yH_2O \rightarrow CaSO_3 \cdot xCaSO_4 \cdot yH_2O \qquad (2\text{-}17)$$

The precipitation reactions shown above must occur if the SO_2 removed from the flue gas is to be discarded as calcium sulfite and calcium sulfate solids. Within certain limits, promotion of these reactions to effect removal of dissolved sulfur from the system is desirable. These precipitation rates must be controlled if the formation of solid scale in the system is to be avoided.

2.1.3 Control of Sulfite and Sulfate Scaling

One of the principle problems encountered in operating lime scrubbing equipment for SO_2 control has been the prevention of scaling. Scaling conditions occur when scrubbers are operated in a mode where calcium, sulfite, and sulfate concentrations are allowed to exceed certain critical levels (OT-023). These levels correspond to critical precipitation rates which, if exceeded, cause scaling of the equipment. These allowable levels may be expressed in terms of the "relative saturations" of the precipitating solids. Relative saturation is a measure of the extent to which the concentration of a precipitating species exceeds the equilibrium solubility of the solid that is formed.

In tests conducted at TVA's Shawnee facility, EPA has confirmed that the maximum allowable saturation level of calcium sulfate is 1.4 times its solubility product. No quantitative level has been established for calcium sulfite, but it is known to be considerably higher than the sulfate limit (EP-008).

As mentioned previously, the level of sulfite oxidation in a scrubbing system affects the scaling potential by setting the requirement for sulfate precipitation (and removal) from the system. The level of oxidation at Paddy's Run is relatively low, i.e., about 5 percent. In addition, there is evidence of sulfate-sulfite co-precipitation (BO-147). The combination of these factors allows the Paddy's Run No. 6 system to run in a mode whereby the recirculating liquor is subsaturated in calcium sulfate, thereby eliminating the sulfate scaling problem.

The calcium sulfite scaling is controlled by two different but related phenomena. First, the scrubber feed liquor supersaturation is required to be low enough so that the additional SO_2 absorbed in the scrubber does not produce a liquor which exceeds the critical supersaturation. The scrubber feed supersaturation is controlled by providing sufficient residence time in the hold tank and by recirculating large amounts of sulfite seed crystals to provide growth sites for precipitating solids. At LG&E, a 10% solids slurry is circulated to provide these crystals. This combination of a large surface area for mass transfer (precipitation) and a long residence time makes it possible to sustain required precipitation rates at low driving force (sulfite supersaturation) levels.

The second factor which can contribute to scaling is related to the mechanism by which the scrubber pH is controlled. This can best be explained in terms of the process chemistry. Lime dissolution produces an increased concentration of hydroxyl ions (OH^-) and calcium ions (Ca^{++}):

$$Ca(OH)_2(s) \rightarrow Ca^{++} + 2(OH^-) \qquad (2\text{-}18)$$

The scrubbing liquor contains a large amount of dissolved SO_2 in the form of the bisulfite ion (HSO_3^-). The increased basicity produces a shift in the bisulfite ion equilibrium

$$OH^- + HSO_3^- \rightarrow SO_3^= + H_2O \qquad (2\text{-}19)$$

The resulting increase in the sulfite ion concentration combined with the increased calcium ion concentration increases the driving force for calcium sulfite formation.

$$Ca^{++} + SO_3^= + \tfrac{1}{2}H_2O \rightarrow CaSO_3 \cdot \tfrac{1}{2}H_2O(s) \qquad (2\text{-}20)$$

Again, if the critical sulfite supersaturation is exceeded, scaling results. The LG&E system avoids this phenomenon by controlling the pH of the scrubber feed liquor at approximately 8. At this pH most of the lime which is added to the system dissolves in the hold tank.

2.2 Functional Description

The Paddy's Run No. 6 SO_2 lime scrubber is installed on an existing 65 Mw. boiler, which is used primarily to provide power for peak load periods. The SO_2 scrubbing system consists of two marble bed scrubbers (each with two marble beds), a hold tank (reaction tank), a reaction surge tank, a clarifier, and two vacuum filters. A schematic flow diagram of the system is shown in Figure 2-1. (Only one scrubber is depicted in the diagram.) Typical operating volumes for the various tanks are shown on the flowsheet. Flow parameters for the important streams are listed in Table 2-1.

2.2.1 Flue Gas Introduction

The flue gas from the boiler passes through a 95% efficient electrostatic precipitator. The gas enters a duct arrangement which allows an infinite turndown ratio for the system. Normally only a 20% reduction in flue gas flow to the scrubber can be tolerated without affecting scrubbing efficiency. However, opening the by-pass dampers (see Figure 2-1) allows scrubbed flue gas to be recirculated, thus providing a means for maintaining proper gas flow to the scrubber over a wide range of boiler loads. The price paid for this turndown capability is the power required for the recirculation of gas and the additional reheat requirement. This recirculation mode of operation is used when changing from half to full load or

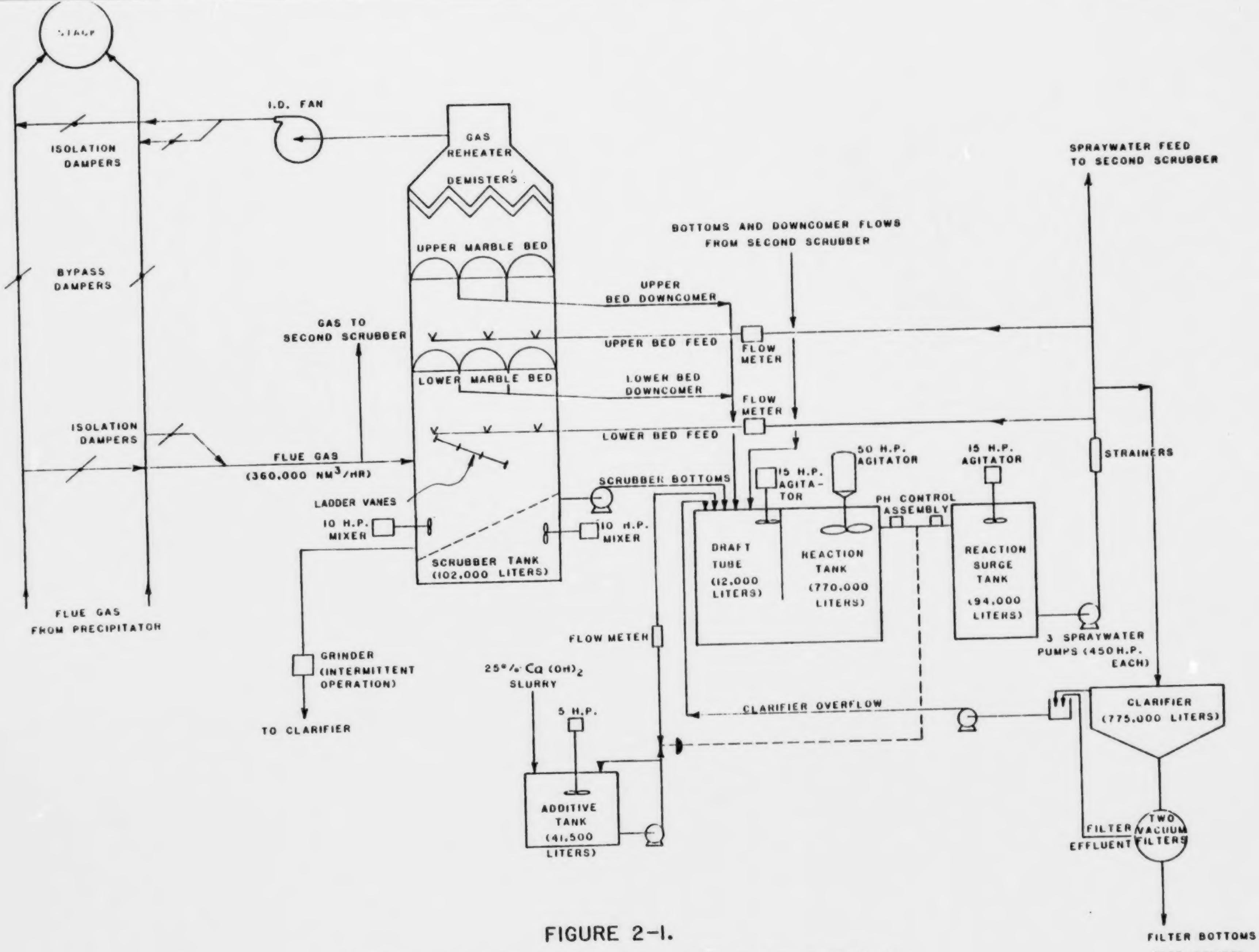

STACK
I.D. FAN
ISOLATION DAMPERS
BYPASS DAMPERS
ISOLATION DAMPERS
FLUE GAS FROM PRECIPITATOR
GAS REHEATER
DEMISTERS
UPPER MARBLE BED
LOWER MARBLE BED
GAS TO SECOND SCRUBBER
FLUE GAS (360,000 NM³/HR)
LADDER VANES
10 H.P. MIXER
SCRUBBER TANK (102,000 LITERS)
10 H.P. MIXER
GRINDER (INTERMITTENT OPERATION)
TO CLARIFIER
UPPER BED DOWNCOMER
UPPER BED FEED
LOWER BED DOWNCOMER
LOWER BED FEED
SCRUBBER BOTTOMS
FLOW METER
FLOW METER
BOTTOMS AND DOWNCOMER FLOWS FROM SECOND SCRUBBER
15 H.P. AGITATOR
50 H.P. AGITATOR
15 H.P. AGITATOR
PH CONTROL ASSEMBLY
DRAFT TUBE (12,000 LITERS)
REACTION TANK (770,000 LITERS)
REACTION SURGE TANK (94,000 LITERS)
FLOW METER
25% Ca (OH)2 SLURRY
5 H.P.
ADDITIVE TANK (41,500 LITERS)
CLARIFIER OVERFLOW
SPRAYWATER FEED TO SECOND SCRUBBER
STRAINERS
3 SPRAYWATER PUMPS (450 H.P. EACH)
CLARIFIER (775,000 LITERS)
FILTER EFFLUENT
TWO VACUUM FILTERS
FILTER BOTTOMS (40% SOLIDS)

FIGURE 2-1.

PADDY'S RUN NO. 6 SO_2 SCRUBBER FLOW DIAGRAM

TABLE 2-1

TYPICAL FLOW RATES FOR THE LG & E SO_2 SCRUBBING SYSTEM

Stream	Flow Rate	Solids Concentration %
Flue Gas	360,000 Nm^3/hr	-
Upper Bed Scrubber Feed	11,340 ℓ/min	10
Lower Bed Scrubber Feed	11,340	10
Upper Bed Downcomer	7,560(1)	10
Lower Bed Downcomer	11,340(1)	10
Scrubber Drain	3,780(1)	10
Clarifier Feed	470	10
Clarifier Overflow	380	0
Clarifier Underflow	230	20
Filter Bottoms	95	40
Additive Feed	110	25

---

(1) Estimated flows

when operating at intermediate levels. At half or at full load the by-pass dampers remain closed with either one or both scrubbers in service.

The flue gas flow at full load is approximately 360,000 Nm^3/hr (350,000 cfm at 355°F). The sulfur content of the coal burned is generally around 2.5%. This yields typical SO_2 flue gas concentrations in the 2,000 ppm range. Air leakage around the preheater causes the oxygen concentration in the flue gas to be in the 6 to 9 mole percent range. The flue gas enters the scrubber below the lower marble bed where a 90° change in the direction of gas flow is accomplished. Uniform gas distribution below the lower marble bed is obtained through the use of ladder vanes which are washed with water periodically to prevent the buildup of solids (ash and dried slurry). The gas is reheated after leaving the scrubber before being discharged to the stack. The total gas pressure drop across the scrubbing system is from 38 to 46 cm (15 to 18 inches) of water. An induced draft fan returns the gas to the stack and maintains .51 cm (.2 inches) of H_2O pressure differential between the incoming and outgoing gas ducts. Dupont SO_2 analyzers monitor SO_2 concentrations in the inlet and outlet ducts.

2.2.2 Scrubbers

The scrubber system consists of two vertical scrubbers in parallel, each having two marble beds in series. The scrubbers are constructed of carbon steel with a 1.27 cm (0.50 inch) flake lining (fiber-glass-epoxy coating). The marble beds have 304 stainless steel support plates with 7.6 centimeters (3 inches) of marbles on each bed. The spacing between the beds is 2.13 m (7 ft). The scrubber tank, which is located in the bottom portion of the scrubber shell (see Figure 2-1), is constructed of unlined carbon steel.

The scrubbing slurry is introduced through 108 nozzles mounted on distribution headers located .61 m (2 ft) below each bed. The nozzles are made of plastic with a Nordel lining. (These replaced the original stainless steel nozzles which eroded.) The feed lines outside of the scrubber are made of carbon steel, while the lines inside the scrubber are made of stainless steel coated on the external surfaces with fiberglass.

The slurry feed flow rate is maintained at 11,340 liters/min (3,000 gpm) per bed. This corresponds to a liquid to gas ratio (L/G) of 7.6 liters/m^3 (35 to 40 gpm/1,000 acfm). The scrubber feed pH set point is 8. With this L/G and scrubber feed liquor pH, SO_2 removal efficiencies of 80 to 90% are achieved at LG & E.

The lower bed scrubber feed, which is introduced below the bed, contacts the incoming flue gas and adiabatically saturates the gas before it enters the marble bed. The gas carries the slurry onto the bed where intimate gas-liquid mixing occurs. The mixing creates a turbulent layer above the bed where SO_2, CO_2, and oxygen are absorbed by the scrubbing liquor. The pH of the liquor on this bed is typically around 5. The turbulent layer extends 25 cm (10 inches) above the bed. This height is set by the positions of 64 small downcomers on each bed. These combine to form a single large downcomer. All downcomer lines are constructed of fiberglass. Approximately 11,340 liters/min (3,000 gpm) of spent liquor are removed from the lower bed by gravity flow through these downcomers. The slurry residence time on both marble beds is approximately 30 to 40 seconds.

An estimated 3,780 liters/min (1,000 gpm) of the 11,340 liters/min (3,000 gpm) scrubber feed liquor is not carried into the lower marble bed. This "weep" (bed reject) liquor collects in the scrubber bottoms tank (102,000 liters). A sloping carbon steel screen prohibits any large solid particles from entering the tank. Periodically, these are removed from the screen, ground in a comminutor, and pumped to the reaction tank. The slurry which passes through the screen and collects in the scrubbing tank is pumped to the reaction tank by the scrubber drain pumps. There are three rubber lined scrubber drain pumps, all of which can be connected to either scrubber. Generally, only one is used per scrubber, with one remaining as a spare. The level in the scrubber tank is maintained by a level controller to help protect the carbon steel tank and screen from corrosion. To prevent settling, each scrubber tank is mixed by two 10 horsepower 316 stainless steel agitators.

As the flue gas leaves the lower bed, it disengages from the lower bed liquor and rises until it contacts the upper marble bed feed liquor. The superficial gas velocity at this point is about 2.2 m/sec (7 fps). Again, the gas entrains the scrubbing slurry and carries it onto the upper bed where turbulence facilitates vapor-liquid mass transfer. An estimated 7,560 liters/min (2,000 gpm) of spent scrubbing slurry are drained by gravity from the upper bed. The other 3,780 l/min (1,000 gpm) are rejected to the lower bed. The downcomers from the upper and lower beds combined to form a common downcomer, a 61 cm (24 inch) fiberglass drain line, which carries the spent scrubbing liquor to the reaction tank by gravity flow.

The pH on the upper marble bed is only slightly higher than the lower bed (5.3 as compared to 5). Previous work performed at Combustion Engineering's Windsor facility indicates that SO_2 removal efficiency is proportional to the pH of the liquor on the marble bed. This indicates that the gas-liquid mass transfer is at least partially controlled by the liquid phase resistance. The fact that upper marble bed pH and the lower bed pH are almost the same would result in the beds having equivalent fractional removal efficiencies.

The flue gas which leaves the upper marble bed contacts two sets of fiberglass Chevron demisters where entrained slurry is removed. The lower bank of demisters positioned 1.27 m (4 ft 2 inches) above the upper marble bed support plate. Once every eight hours the demisters are washed with 756 l/min (200 gpm) of make-up water for about ten minutes. Any slurry that has accumulated on the demisters is removed by the washing and returned to the system. The 8 spray wash lines are retractable.

After the gas passes through the demisters, it enters the reheat section. In each scrubber, there are two gas fired igniters positioned about 1.22 m (4 ft) above the demisters. The reheated gas proceeds through the induced draft fan to the stack.

2.2.3 Reaction-Hold Tank

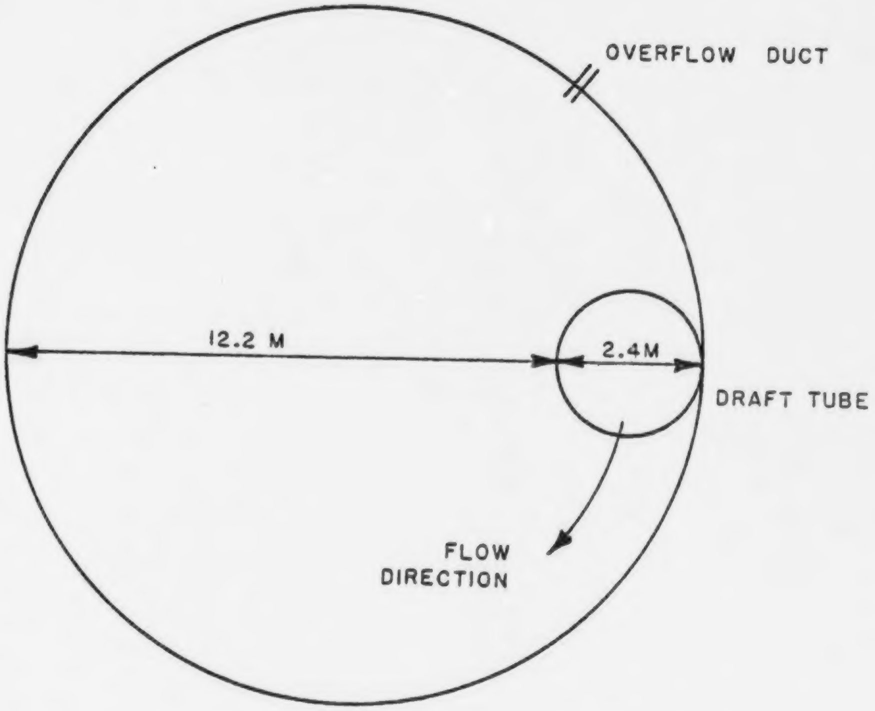
The hold tank is a 770,000 liter (205,000 gallon) capacity stirred tank which is 14.6 m (48 ft) in diameter. To one side within this tank a small draft tube is mounted. The draft tube (12,000 liters) is a 2.4 meter (8 foot) diameter pipe a portion of which is cut-away about 1.8 meters (~6 ft) above the base of the hold tank. Details of this draft tube/hold tank arrangement are shown in Figure 2-2. All streams entering the hold tank are introduced through the draft tube. These feed streams are (1) the combined downcomers from the marble beds, (2) the scrubber bottoms, (3) clarifier overflow, and (4) additive slurry feed. The feed streams are mixed by a 15 horsepower carbon steel agitator and exit through the hole in the bottom of the draft tube. The residence time in the draft tube is approximately 15 seconds.

The combination of the sulfite-rich, spent scrubbing liquor and the high pH lime slurry creates a large driving force for precipitation. As a result, some nucleation probably occurs in this draft tube. The small sulfite crystals which are probably formed in the draft tube act as seed crystals for further precipitation in the hold tank.

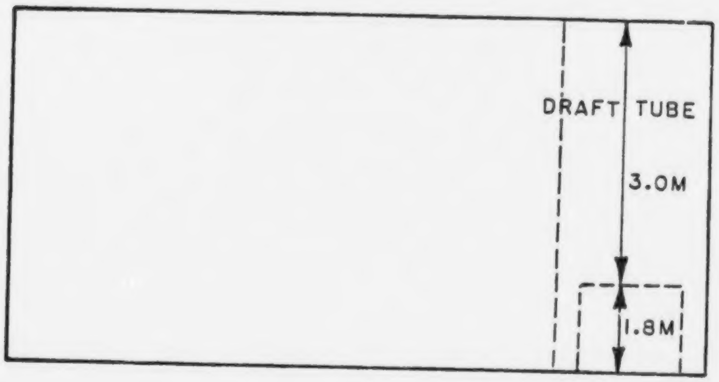
The function of the hold tank is to provide time for the calcium sulfate and sulfite precipitation and calcium hydroxide dissolution reactions to take place. A 50 horsepower, carbon steel agitator is used to keep the slurry mixed. The slurry residence time in the hold tank is approximately 18 minutes when both scrubbers are in operation. Both the seed crystals generated in the draft tube and existing solids particles act as nucleation sites for calcium sulfite precipitation. The residence time of the slurry in the hold tank is sufficient to allow precipitation to occur with corresponding reduction in supersaturation. The hold tank also provides time for the majority of the lime dissolution to occur.

2.2.4 Reaction Surge Tank

The reaction surge tank is fed continuously from the reaction tank by gravity overflow. A pH control assembly in the overflow line adjusts the lime additive flow rate to maintain a


OVERFLOW DUCT
12.2 M
2.4M
DRAFT TUBE
FLOW
DIRECTION

HOLD TANK TOP VIEW


DRAFT TUBE
3.0M
1.8M

HOLD TANK SIDE VIEW

FIGURE 2-2 HOLD TANK - DRAFT TUBE CONFIGURATION

specified pH. The set point is normally at pH 8 to accommodate any SO_2 concentration fluctuations in the flue gas. The system performance is more sensitive to flue gas SO_2 level fluctuations when the scrubber feed liquor is at pH 9.0 or 9.5. The surge tank is mixed by a 15 horsepower, carbon steel agitator. The surge tank provides an additional two minutes residence time for the slurry and dampens out any level fluctuations that might occur in the hold tank.

The slurry from the surge tank is returned to the scrubbers as feed by three 450 hp Nihard centrifugal pumps; one for each scrubber and a spare. When both scrubbers are in operation 45,360 ℓ/min (12,000 gallons per minute) are pumped from the surge tank. The slurry passes through two strainers before entering the scrubbers. A slip stream (~470 ℓ/min) from the reaction surge tank to the clarifier purges solids from the system at the same rate that SO_2 is removed from the flue gas (at steady state), thereby controlling the amount of solids in the system. A 10% solids level in the recirculating slurry is maintained.

2.2.5 Clarifier and Filters

The clarifier is a 775,000 liter (205,000 gallon) carbon steel settling tank in which waste solids are accumulated and concentrated prior to being purged from the system. The retention time in the clarifier is approximately 36 hours. As the solids settle, the clear liquid passes through an overflow weir to a small surge tank. The liquor is pumped from this tank to the draft tube by a 5 horsepower centrifugal pump. The clarifier bottoms (~20% solids) are removed through two lines, 6.35 cm (2½ inches) and 10.2 cm (4 inches) in diameter, and pumped to the filter system by two 5 hp centrifugal pumps.

The filter system consists of two vacuum drum filters. The solid cake (~40% solids) which is formed as water is drawn through the filter, is removed by blades as the drums turn. The filter effluent is returned to the overflow weir of the clarifier where it is combined with the clarifier overflow and returned to the reaction tank. The 100 ℓ/min filter cake is removed from the filter drum and placed on a conveyor. Liquid associated with the filter cake is the only source of blowdown liquor leaving the system. Other than this, the system is completely closed loop.

Shortly after beginning operation of the scrubbing system, it was apparent that the clarifier was not large enough to maintain a 10% solids feed slurry. Due to the low oxidation level at LG&E, the solid crystals formed are primarily calcium sulfite rather than a mixture of calcium sulfite and sulfate. This is due to the low oxidation level. Calcium sulfite settles more slowly than does calcium sulfate. The low oxidation, therefore, helps reduce the chance of scaling but hinders the settling of the solids. Two flocculants, Dow A-23 and Betz 1100, were tested for improvement of the solids settling in the clarifier. Addition of either of the flocculants improved the solids setting, but use of the Betz 1100 flocculant improved the settling more.

2.2.6 Lime Addition

The alkali used at LG&E is carbide lime, a by-product of a nearby acetylene manufacturing operation. An analysis of this material is shown in Table 2-2. The lime is delivered by truck from large storage piles as a 25% solid slurry. This is fed through a disintegrator which reduces the size of any large particles and removes most of the ferrous materials which may be present. The lime slurry passes to the 41,600 liter (11,000 gallon) additive slurry tank.

TABLE 2-2

CARBIDE LIME ANALYSIS

Components	Wt. %
Calcium hydroxide - $Ca(OH_2)$	92.50
Calcium carbonate - $CaCO_3$	1.85
Silica - SiO_2	1.50
Iron and alumina oxides - R_2O_3	1.60
Magnesium oxide - MgO	0.07
Sulfur	0.15
Phosphorus	0.01
Free carbon	0.25
Not analyzed	2.07

Source: Airco Carbide Lime Technical Data and Process Bulletin, April, 1969.

The additive tank is kept well mixed by a 5 hp carbon steel agitator. As mentioned previously, the lime slurry is fed to the reaction tank on pH demand. Typically, lime is added at a rate which maintains a hold tank effluent pH of 8. Roughly 115 liters/min (30 gpm) of 25% lime slurry is utilized when both scrubbers are operating. A 10 horsepower Nihard pump is used in this system. An additive recirculation line is utilized to facilitate mixing and to maintain velocities in the additive feed loop over the full range of operation. As the control valve restricts flow to the reaction tank, the amount of recirculation increases. To prevent plugging while operating at low loads, the control valve is opened when the slurry flow meter indicates low flow. The level in the additive tank is monitored by a level indicator.

2.2.7 Water Addition

When the system is operating at full load, roughly 455 liters/min of H_2O is vaporized in the scrubber and exits with the stack gas. In addition, another 57 liters/min of water leaves the system with the filter cake. The majority of the make-up water is introduced with the lime slurry feed and injected into the packing glands on the pumps. In addition, water is introduced into the system as demister wash water and ladder vane spray wash water. An accurate record of the water input has not been kept and the flow is somewhat variable due to fluctuations in the water flow to the packing glands.

3.0 OPERATING HISTORY

In this section, the general operating history of the LG & E Paddy's Run No. 6 lime scrubber is presented. Unit availabilities and SO_2 removal efficiencies are provided. Process and equipment problems which were encountered during operation and their solutions are also discussed.

3.1 General History

The lime scrubbing system at Paddy's Run No. 6 was started up in April, 1973. Through August, 1974 "Scrubber A" had been operated 1,818 hours and "Scrubber B" 3,015 hours. Paddy's Run is used only during peak load periods. For this reason it operates at reduced loads quite frequently. During those periods of time when the boiler was operating at or less than half load and only one scrubber was operated, the "B" scrubber was generally used. This accounts for the difference in operating time for the two scrubbers.

From April until August of 1973, general start-up problems were encountered and system experimentation was conducted. The vast majority of the operating time occurred during the last four months of 1973, during 20 days in July and August, 1974, and during October, 1974. The period from September to December, 1973, was used to demonstrate system performance and reliability. In October and November, the system was operated continuously for 35 days and 14 hours. This was the longest period of continuous operation logged to date. During July - August, 1974. (a peak load period) the Paddy's Run No. 6 unit was operated at full load during a substantial portion of the period. As a result a considerable amount of scrubber operating time was logged during this period. The magnesium content of the lime used at this time was higher than the magnesium levels of the lime which had been utilized previously. The October, 1974, run was a special limestone test for which Combustion Engineering contracted the Paddy's Run No. 6 scrubbing system. A summary of the operating history is shown in Figure 3-1. The important system parameters for October and November, 1973, and for July, 1974 are plotted in Figures 3-2 to 3-10. These periods represent significant results for the low magnesium and high magnesium lime operation respectively.

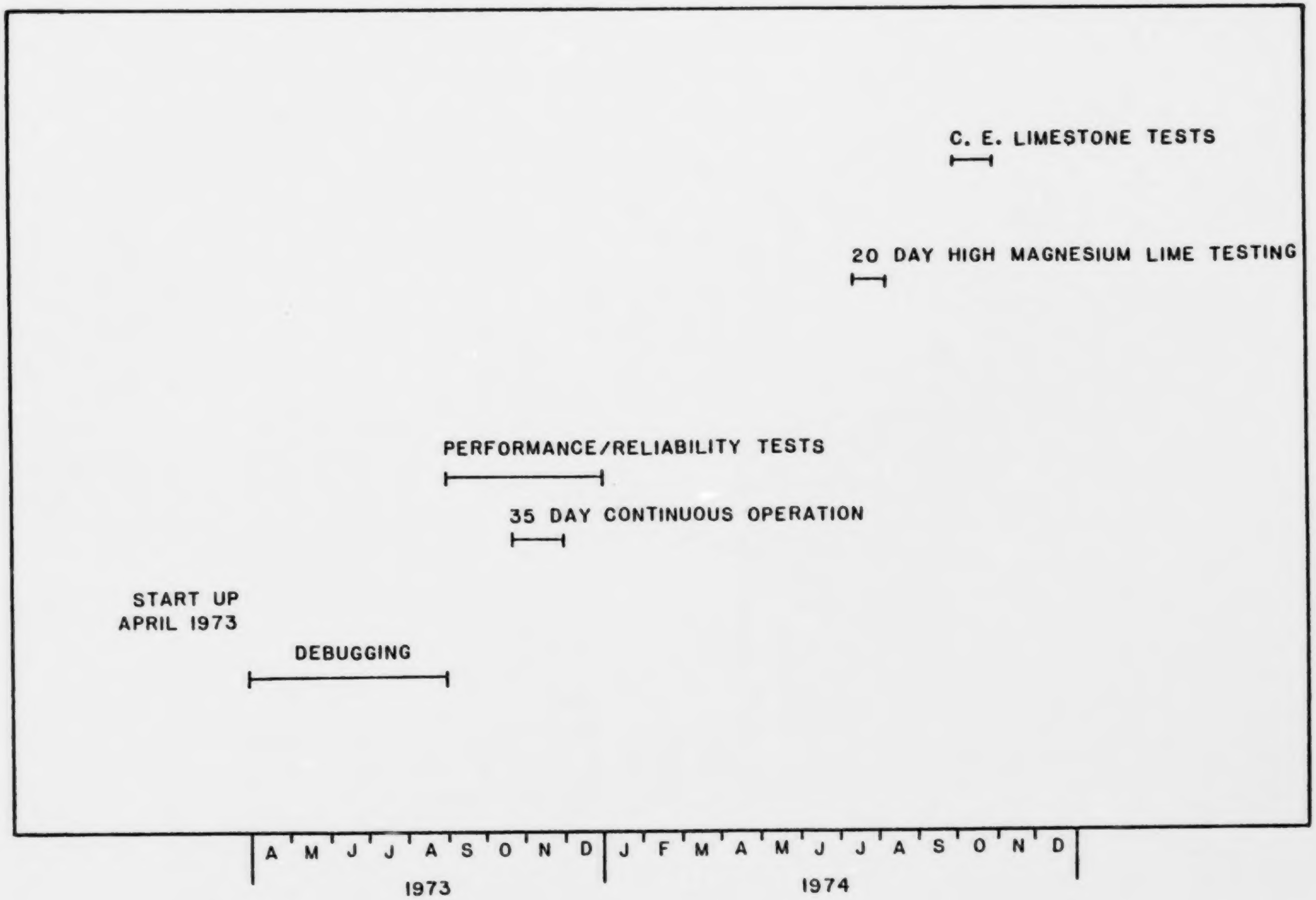
The parameters which were plotted include hours of operation for Scrubbers A and B, inlet SO_2 concentrations and daily SO_2 removal efficiencies for both scrubbers. The scrubber feed and effluent stream pH's were also recorded. (The scrubber effluent pH shown is the average pH for the Scrubber A and B lower bed downcomers. The pH's of these streams always agreed to within 0.2 pH units. In addition, the lower bed downcomer stream was felt to be representative of the other effluent streams since there were never more than 0.5 pH units difference among the upper bed downcomer, the lower bed downcomer, and the scrubber bottoms streams.) The solids content of the circulating slurry and the liquid species composition of the hold tank effluent (or scrubber feed stream) are also shown as a function of time. A daily maximum and minimum for the parameters are indicated where information was available and meaningful.

The period in which the system operated with higher magnesium content lime had significantly higher SO_2 removal efficiencies than did the period during which lower magnesium lime was utilized. Typical magnesium concentrations in the October - November, 1973, run period were about 40 ppm while the July, 1974, levels were in the 1000 to 2000 ppm range. Removal efficiencies of 80 to 90% were reported for the October - November, 1973, operation. All of the removal efficiencies for July, 1974, were above 95%.

FIGURE 3-1

PADDY'S RUN NO.6 FLUE GAS DESULFURIZATION

SYSTEM OPERATING SUMMARY



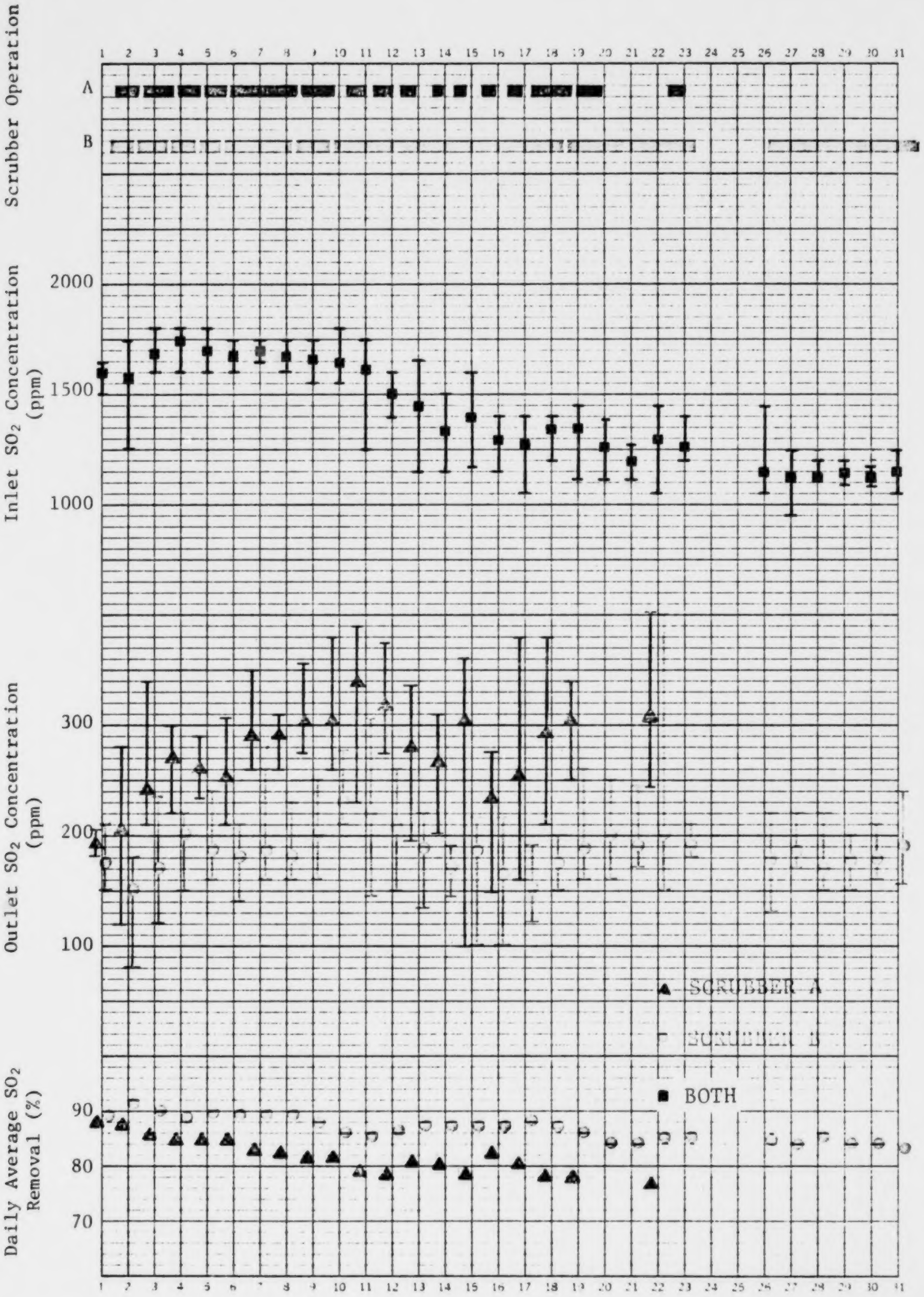


FIGURE 3-2 LG&E OPERATING LOG AND SO_2 REMOVAL FOR OCTOBER, 1973

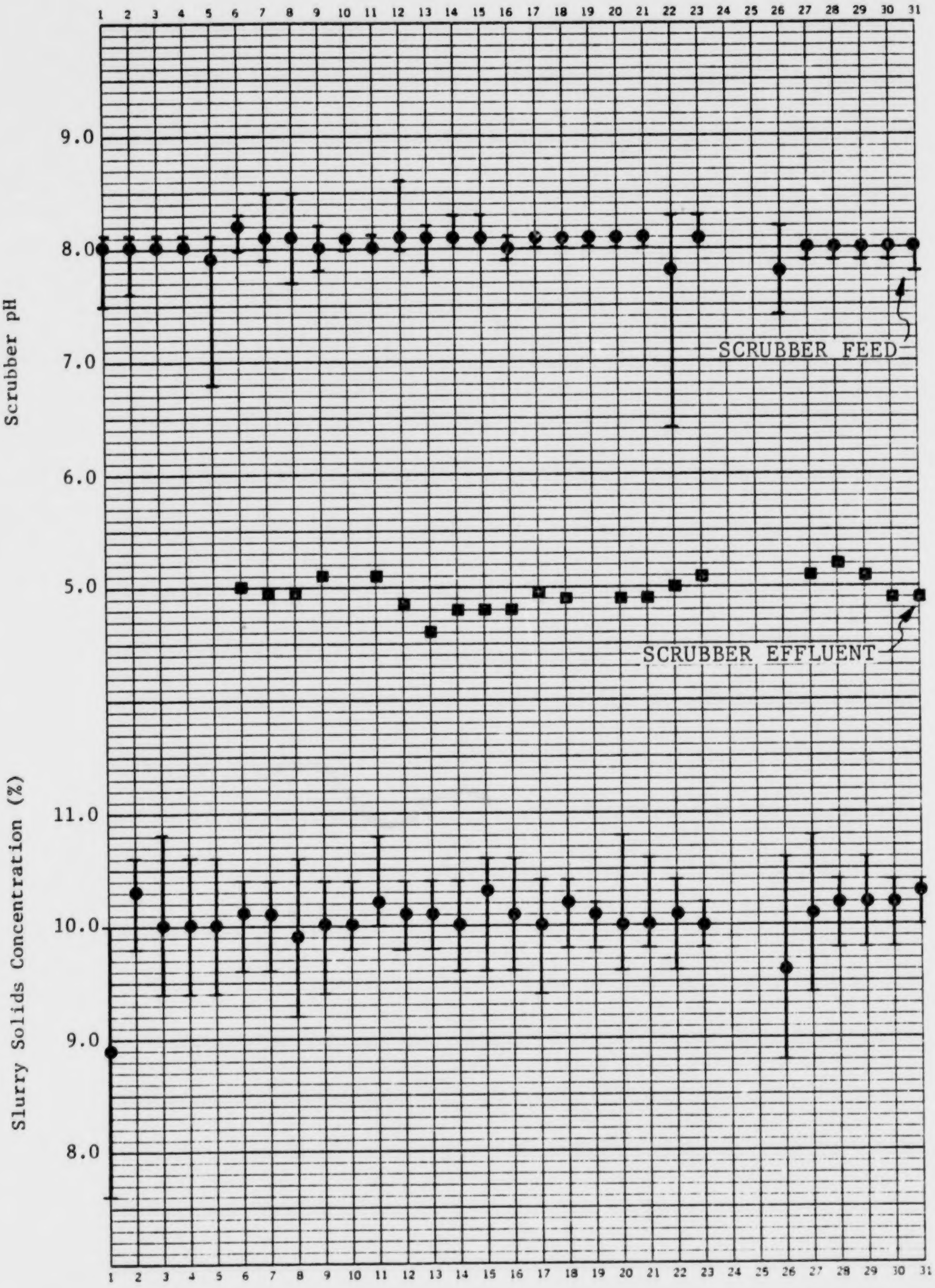


FIGURE 3-3 LG&E SCRUBBING LIQUOR PARAMETERS FOR OCTOBER, 1973

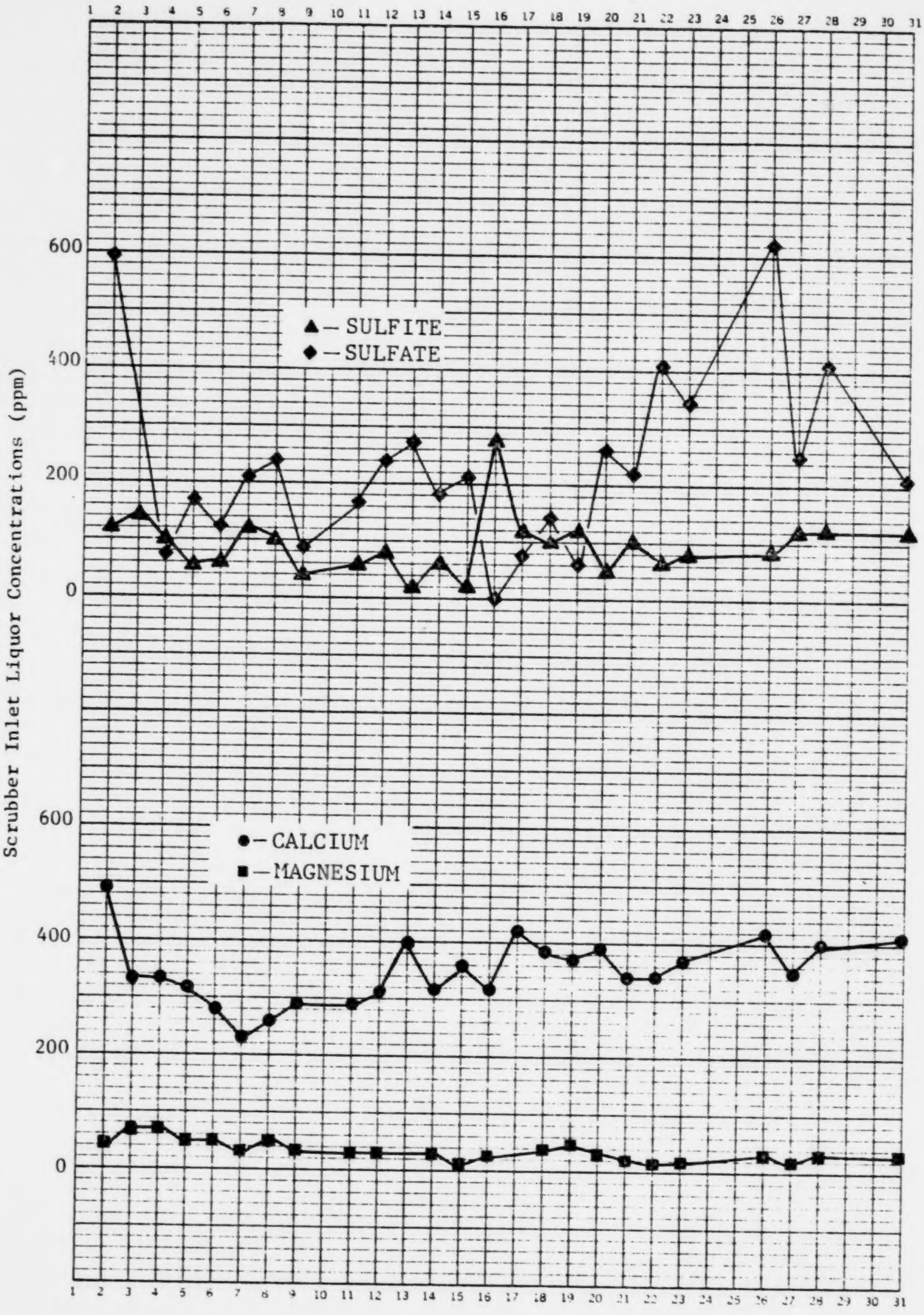


FIGURE 3-4 LG&E DISSOLVED SOLIDS CONCENTRATIONS FOR OCTOBER, 1973

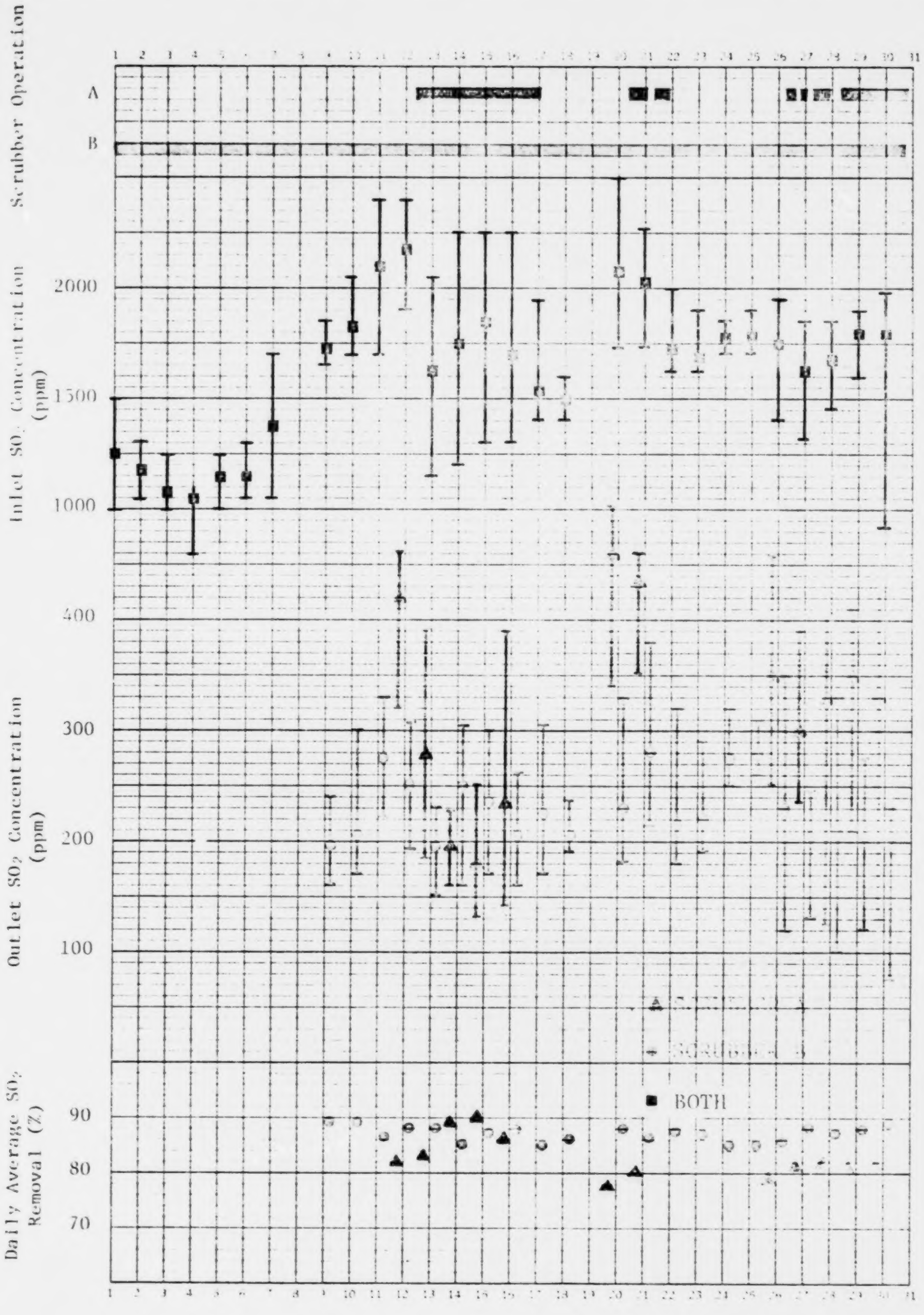


FIGURE 3-5 LG&E OPERATING LOG AND SO$_2$ REMOVAL FOR NOVEMBER, 1973

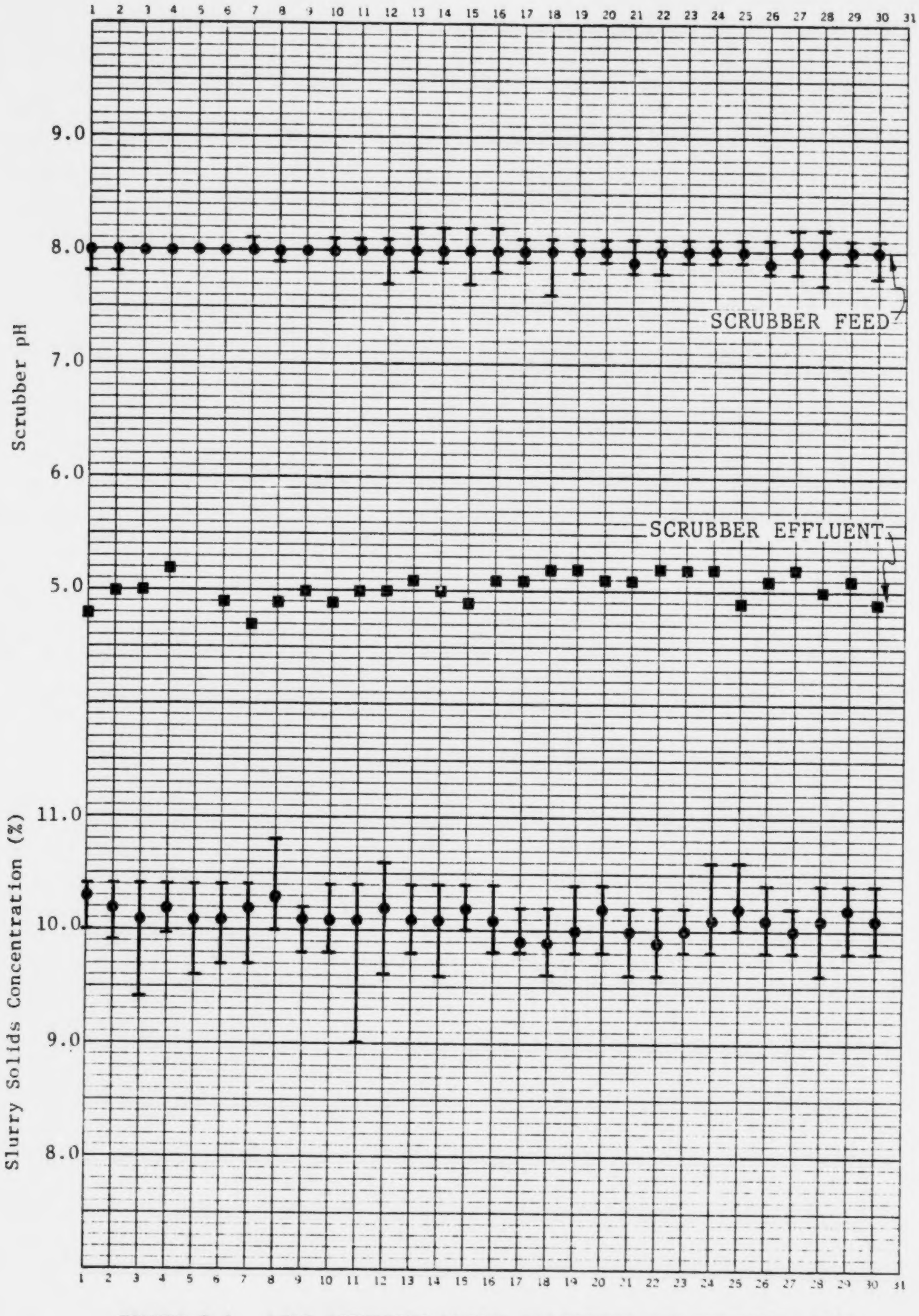


FIGURE 3-6 LG&E SCRUBBING LIQUOR PARAMETERS FOR NOVEMBER, 1973

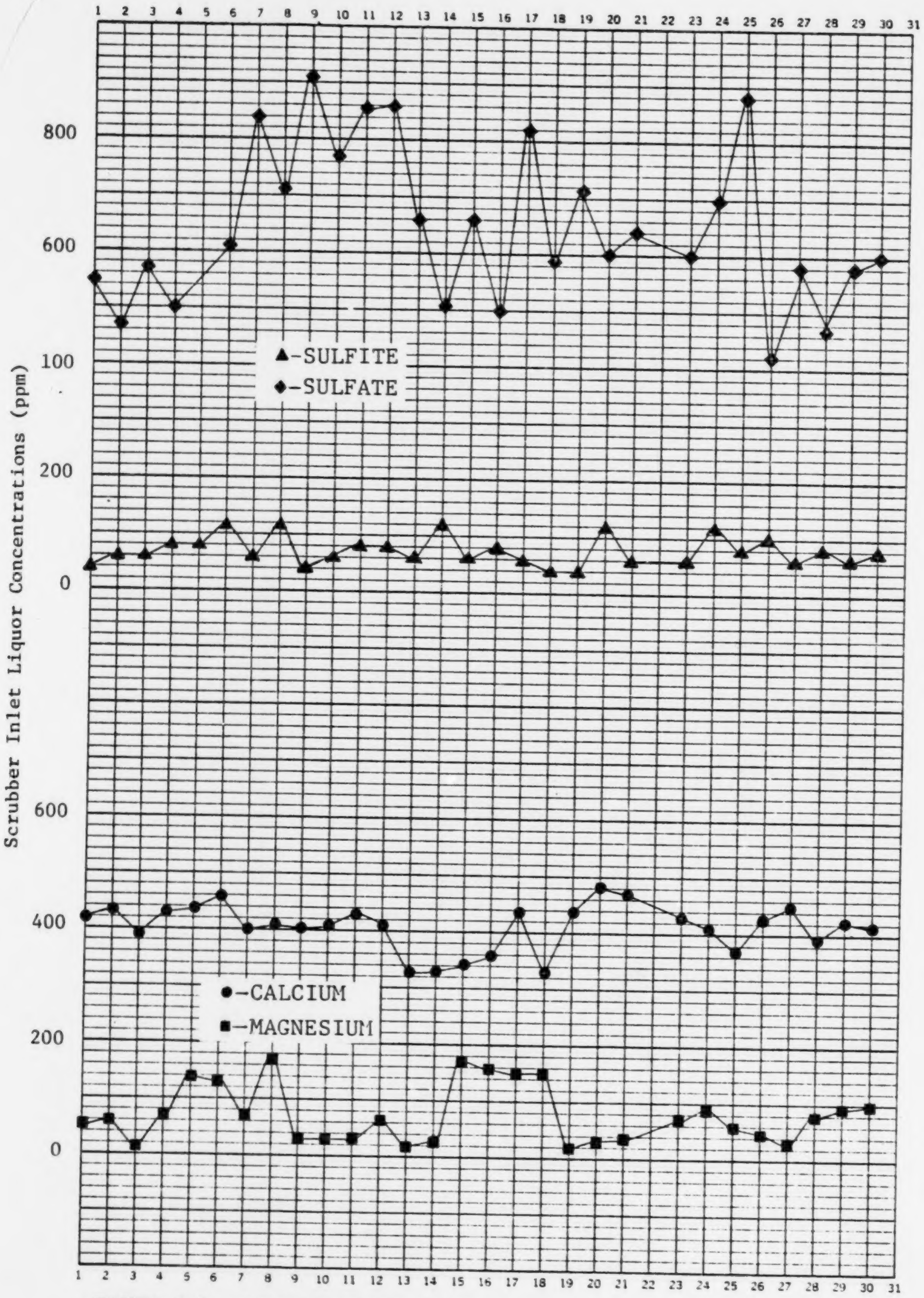


FIGURE 3-7 LG&E DISSOLVED SOLIDS CONCENTRATIONS FOR NOVEMBER, 1973

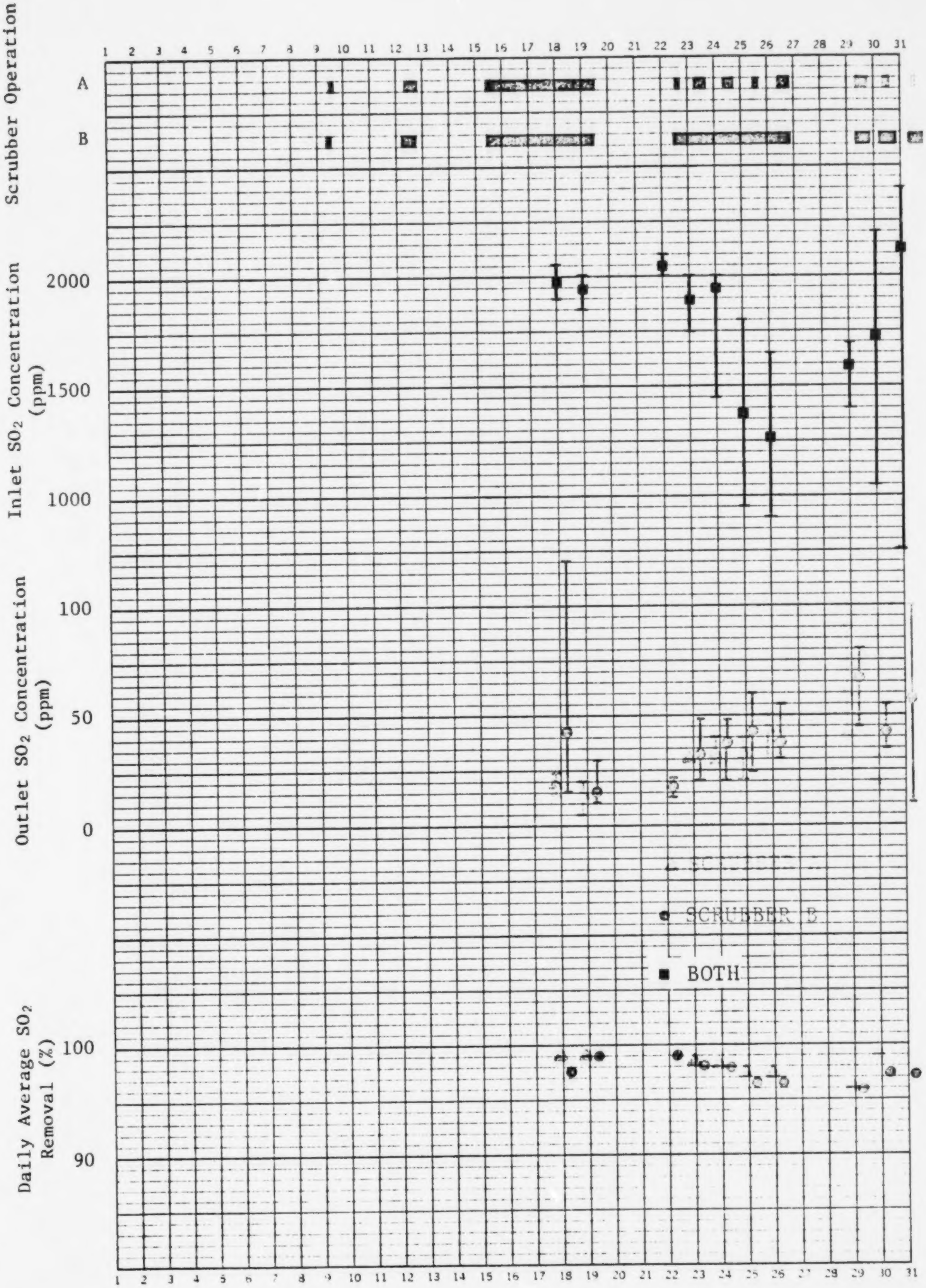


FIGURE 3-8 LG&E OPERATING LOG AND SO_2 REMOVAL FOR JULY, 1974

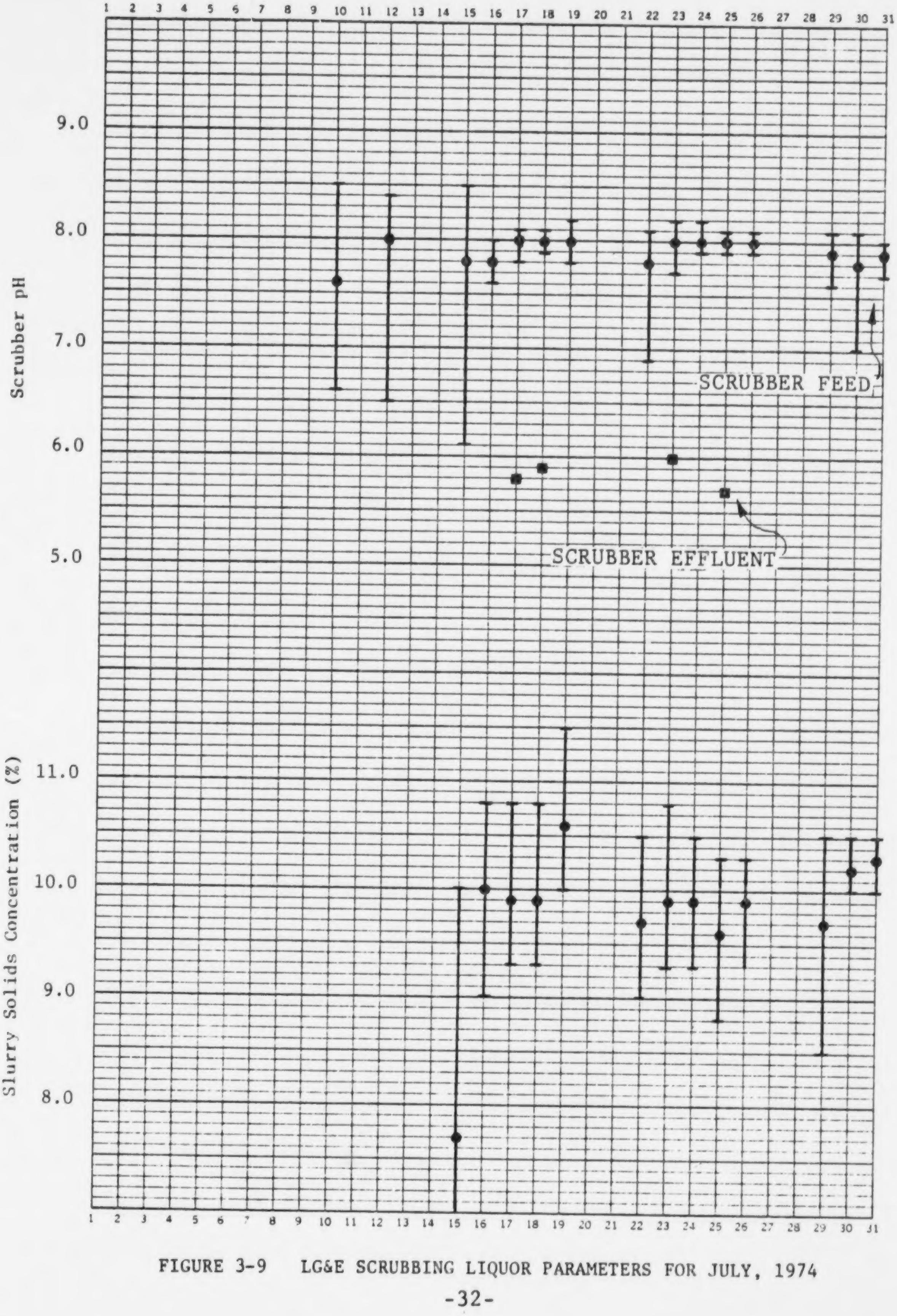


FIGURE 3-9 LG&E SCRUBBING LIQUOR PARAMETERS FOR JULY, 1974

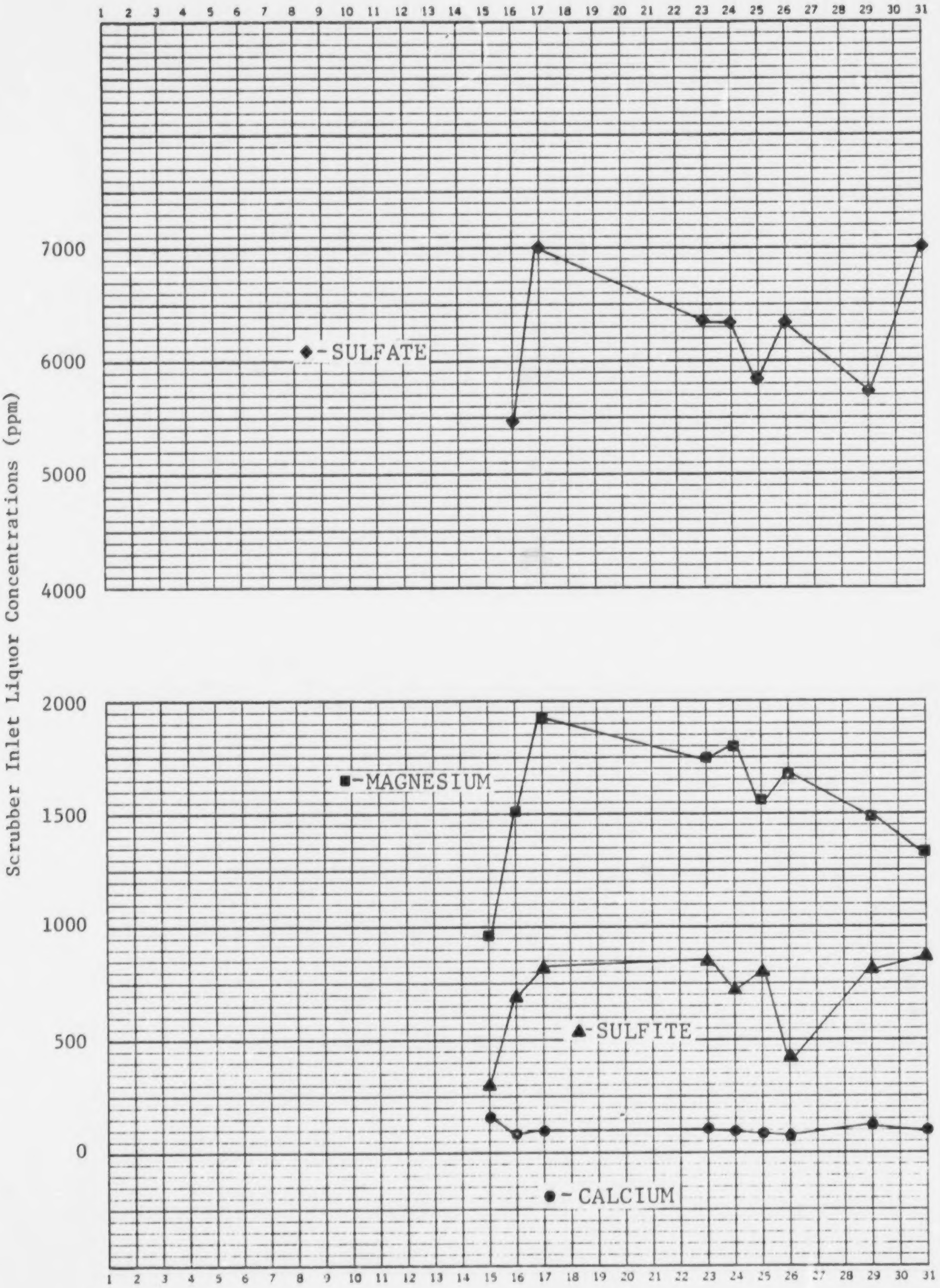


FIGURE 3-10 LG&E DISSOLVED SOLIDS CONCENTRATIONS FOR JULY, 1974

In general, the control of the scrubber feed stream pH and slurry solids concentration was adequate. During the October - November, 1973 period, the daily pH fluctuations were usually within 0.5 pH units. A range of less than one percent was typical of the solids concentration fluctuations. These ranges for July, 1974, were somewhat greater. This was probably due to more sporadic system operation during this period.

The hold tank effluent chemical analyses data support the expected trends with respect to increased magnesium concentrations. For a ten-fold increase in magnesium concentration (from 100 to 1000 + ppm) the dissolved sulfate concentration increased from about 500 ppm to over 6000 ppm. The dissolved sulfite content of the hold tank effluent liquor changed from approximately 100 ppm to over 500 ppm. These trends are supported by laboratory and pilot data (BO-097). However, daily fluctuations in the reported sulfate ion concentration probably indicate some analytical errors. Changes from 300 to 1000 ppm in the sulfate concentration in a day's time are reported on several different occasions. It is not likely that changes in system operations could account for fluctuations of that magnitude over the time frame involved.

Accompanying increases in magnesium, sulfate, and sulfite concentrations is a rise in the pH of the scrubber effluent from about 5 (at low Mg levels) to almost 6. This is probably an indication of the increased scrubbing capacity of the liquor.

3.2 Unit Availability

The unit availabilities which were reported by PEDCO for the Paddy's Run No. 6 lime scrubbing system are shown in Table 3-1 (PO-132). As mentioned previously, the first five months of operation involved solving initial start-up problems as well as experimentation with system procedures. For these reasons, the availability figures for the period of April through August, 1973, are not very meaningful. Although the availability figures for the other operating periods give a better indication of operational reliability, interpretation of these figures presents problems.

Reporting availability figures for scrubber units is particularly difficult when the boiler is only used during peak periods. The definition of availability used by PEDCO is the time of scrubber operation divided by the time of boiler operation. When the Paddy's Run No. 6 boiler operates at half load, only one of the two scrubbers is operated. However, the availability definition used by PEDCO does not account for this. The scrubber which was not operated due to low demand has a reported availability which does not give a true indication of that scrubber's actual availability. Since Scrubber A was usually operated only to satisfy full load demand, the reported availability for this scrubber is usually less than that for Scrubber B and does not give an accurate indication of the Scrubber A's performance.

Perhaps a more appropriate basis for calculating availability figures for the LG&E scrubber system is a computation of the scrubber operating time required to scrub the flue gas produced during boiler operation. The required scrubber operation is twice the time the boilers were at full (two scrubbers were required) plus the time the boiler was operated at half load

TABLE 3-1

LG & E PADDY'S RUN NO. 6 SCRUBBER AVAILABILITY

(as reported by PEDCO)

Year	Month	Availability* Scrubber A	Availability* Scrubber B
1973	April	18	56
	May	11	65
	June	.1	6
	July	21	21
	August	53	64
	September	85	72
	October	49	94
	November	35	100
	December	44	78
1974	July	51	81
	August	50	77
	October	100	100

---

*Availability = $\frac{\text{hrs. scrubber operating}}{\text{hrs. boiler operating}} \times 100$

Source: (PO-132)

(one scrubber was required). This parameter is divided into the total hours of scrubber operation (Scrubber A operating time plus Scrubber B operating time) and multiplied by 100 to calculate the % availability for the system. The availability figures calculated according to this procedure are presented in Table 3-2 for comparison with availability numbers reported by PEDCO.

The availability figures calculated in this manner are generally higher than the PEDCO figures. After August, 1973, the system was available about 95% of the time.

Another problem involved with interpreting reported availability data is connected with the fact that the Paddy's Run power plant is staffed with sufficient personnel to operate only two boilers and the scrubber system. Any time more than two boilers at Paddy's Run are required to run, the scrubber system is not operated. These occasional periods reduce the availability figures even though the scrubbers themselves might be available for operation. Unfortunately these periods were not reported in the scrubber system status reports. There are also times when the boiler was not expected to be operated for long periods and the scrubbers were therefore not started up. These periods also reduce the reported availability.

3.3 SO_2 Removal Efficiency

The SO_2 removal efficiencies for the scrubbing system have typically been 80% to 90%. A summary of these efficiencies is presented in Table 3-3. Sulfur dioxide removal efficiencies from 80% to 90% were achieved with a lime feed having a low magnesium hydroxide content (0.7%). In July, 1974, the efficiency

TABLE 3-2

PADDY'S RUN NO. 6 SCRUBBER SYSTEM AVAILABILITY

Month	Boiler Time At Full Load	Operation Time At Half Load	Scrubber Operation* Required	Total Scrubber** Operation (hrs)	Availability***
1973					
August	188	142	518	388	74.9
September	249	140	638	610	95.6
October	332	340	1,012	975	96.3
November	223	495	941	968	100.0
December	43	149	235	235	100.0
1974					
July	143	100	386	365	94.6
August	11	14	36	29	80.6

---

* Scrubber Operation Required = 2 x Time at Full Load & Time at Half Load

** Total Scrubber Operation = Scrubber A Operating Time & Scrubber B Operating Time

*** Availability = $\frac{\text{Total Scrubber Operation}}{\text{Scrubber Operation Required}} \times 100$

TABLE 3-3

SO_2 REMOVAL EFFICIENCIES FOR PADDY'S RUN NO. 6 LIME SCRUBBERS

Operating Period	Removal Efficiency	
	Scrubber A	Scrubber B
Sept. 2 - 8, 1973	87.7	86.5
Sept. 16 - 22, 1973	86.0	89.3
Sept. 30 - Oct. 6, 1973	86.0	89.4
Oct. 7 - 13, 1973	81.1	87.3
Oct. 14 - 20, 1973	78.9	86.7
Oct. 21 - 27, 1973	74.4	84.4
Oct. 28 - Nov. 3, 1973	-	84.6
Nov. 4 - 10, 1973	-	88.3
Nov. 11 - 17, 1973	88.2	87.2
Nov. 18 - 24, 1973	-	86.1
Nov. 25 - Dec. 1, 1973	-	85.7
Dec. 2 - 8, 1973	-	93.9
Dec. 9 - 15, 1973	88.4	93.5
Dec. 16 - 22, 1973	91.5	94.4
July 14 - 27, 1974	98.3	97.9
July 29 - Aug. 4, 1974	97.9	96.8
Aug. 5 - Aug. 11, 1974	96.2	95.9

jumped to 95+%. This increase in scrubbing efficiency was attributed to an increase in the magnesium hydroxide content of the lime feed (to 2.2%).

3.4 Process Problems

The Louisville Gas and Electric scrubbing system at Paddy's Run No. 6 has run for approximately six months during the past two years with few process problems. The major process problems encountered and the solutions which were attempted are discussed in this subsection.

3.4.1 System Turndown

One of the first problems noted at LG&E involved the system turndown ratio. Since Paddy's Run No. 6 is less efficient than many other units in Louisville Gas and Electric's power system it is used in peaking service. For this reason, it is frequently run at less than full load when it operates. However, at load levels less than 80% of maximum load, the SO_2 scrubbing efficiency of the marble beds was reduced. At levels between 50% and 80% of maximum load, the gas velocity was too low for dual scrubber operation and too high for operation of a single scrubber.

This problem was alleviated by flue gas recirculation. A system of baffles in the flue gas ducts was installed so that gas could be recirculated when the boiler was operating between half and full load. As shown in Figure 2-1, an induced draft fan is used to maintain a pressure differential between the outlet and inlet ducts. When operating at less than design gas flows with the bypass dampers and the isolation dampers open, the higher pressure in the outlet duct will force some of the cleaned gas to be recirculated back to the inlet. Recirculation gas flow will increase to the point that the scrubber gas phase pressure drop is equal to the ID fan ΔP. This occurs, of course, when scrubber vapor flows reach design levels. The gas velocity through the scrubber is maintained at a level which is high enough to maintain good scrubber hydraulic operating characteristics. When the system is operated at half or full load, the bypass dampers are closed and the gas flows through on a once through basis. With gas recirculation, the scrubbing system has infinite turndown capability. The cost of this turndown capability is the increased power the ID fan requires and the increased reheat cost.

3.4.2 Solids Control

Initially, when the system was operated at full load with both scrubbers in service, the solids level in the circulating slurry could not be controlled because of solids in the clarifier overflow. Chemical analysis showed that the vast majority of the solid particles leaving the system were calcium sulfite rather than a mixture of calcium sulfate and calcium sulfite as had been expected (due to the low levels of sulfite oxidation which were being obtained). Since calcium sulfite crystals exhibit poorer settling characteristics than do calcium sulfate crystals, the clarifier at LG&E was greatly undersized. Thus, some means of improving the solids settling characteristics had to be found. Two flocculating agents, Dow A-23 and Betz 1100, were added to the clarifier in separate tests. Better clarification was obtained with the Betz 1100. With the addition of either flocculant, the solids in the scrubbing loop could be maintained at 10%. However, once this solids control problem was solved, it was noticed that the slurry volume was still slowly increasing. It was found that the amount of water added to the system was not being properly monitored, and this excess water make-up was contributing to the slurry volume build-up. Once the amount of water being used was reduced, this problem disappeared.

3.4.3 Clarifier Pluggage

During the attempted extended run in October, 1973, the clarifier underflow line became plugged with a jelly-like substance which was identified as concentrated flocculant. The source of this problem was the procedure for flocculant addition. The flocculant had been added at rates corresponding to full load operation (two scrubbers in service). When the boiler load was reduced by half, only half as much flocculant was required. During the extended run the length of operation at half load was significant. Over a period of time, the excess flocculant settled at the bottom of the clarifier and eventually blocked the underflow line. This problem was solved by simply reducing the amount of flocculant added when the boiler load was reduced.

3.4.4 pH Control

Lime additive is fed to the reaction-hold tank on pH demand. The pH set point is typically eight. When the pH in the reaction tank exit drops below the set point, the control valve in the lime additive line opens allowing more 25% lime slurry to enter the hold tank. During the extended run, pH control problems were traced to a plugged control valve.

Again, the problem was due to a load level change. When the system is operating at reduced load, less lime is required and the control valve closes. This means that more lime slurry is recycled to the lime additive tank through the recirculation line. A certain amount of lime deposits around the reduced aperture in the control valve. After extensive operation at reduced load during the "long" run, this deposit hardened. When the system was returned to full load, the deposit restricted the additive flow which caused the pH in the scrubbing loop to drop.

To prevent this incident from recurring, two changes were made in the procedure for reduced load operation. First, the pH control valve is now opened to allow 100% flow for two seconds once every four hours. This reduces the chances of having deposits accumulate in the valve. Secondly, the valve in the recirculation line is opened two additional turns to reduce the pressure in the line at the control valve when the system is operating at reduced load.

3.4.5 Scaling

The system has remained free of sulfate scaling problems throughout its operation using carbide lime. The demisters have remained clean with the exception of particulate scale caused by an electrostatic precipitator malfunction during May 1973. However, in September, 1973, sulfite scale was noted in the scrubbers, primarily on the marble beds.

The scale was noticed while the boiler was burning low sulfur coal. At this time, the pH control set point for the scrubber feed was 9.5. It was reasoned that the change to low sulfur coal reduced the amount of SO_2 in the flue gas which, in turn, increased the pH of the liquor on the marble beds. The increased pH caused a shift in the sulfite-bisulfite equilibrium with the sulfite ion being the more prevalent species. This caused calcium sulfite precipitation and scaling in the scrubber.

The majority of the scale was removed by lowering the pH of the scrubber feed so that a reverse shift in the sulfite-bisulfite equilibrium occurred. The scale clean-up took approximately five hours. The scale did not completely disappear during the run and the remainder had to be cleaned manually. The pH set point of the scrubber feed was reduced from 9.5 to 8 to more adequately cover fluctuations in sulfur content of the coal. This change continued to provide good SO_2 removal and prevented further sulfite scaling.

3.4.6 Scrubber Tank Slurry Level

There have been problems controlling the liquid level in the system, primarily in the scrubber tank. During early operation, failure to monitor seal water flow to the pumps resulted in a liquid level rise during early operation. After the amount of seal water was reduced to reasonable levels, this problem disappeared. At the end of the extended run in December, 1973, there were several corroded areas noted on the scrubber tank screen. The probable cause for the corrosion was a low tank liquid level which exposed portions of this screen and the sides of the tank to the SO_2-rich flue gas. The level in this tank is critical because the tank is constructed of carbon steel and thus is vulnerable to attack by acidic liquors (which would result from SO_2 sorption by stagnant liquid droplets on scrubber surfaces). The sloping screen was reinforced with stainless steel in the weakened areas and the level has been closely monitored recently to prevent this from recurring.

3.5 Equipment Problems

The Paddy's Run lime scrubbing system has encountered several equipment malfunctions, some of which are directly related to the conditions of service and others which are not. Since there is a 10% solids slurry being circulated, there were a number of erosion problems which appeared after a significant period of operation. This subsection discusses the equipment problems which were caused by the process or which played an important part in system operation. Attempted solutions are discussed.

3.5.1 SO_2 Monitoring

Difficulties with the Dupont SO_2 on-line analyzer developed early in the operation at Paddy's Run and have been a continuing problem. The major trouble appears to be caused by dirt, water, and oil in the air supply. Valves and seals corroded and had to be replaced. Frequently, lack of confidence in the analyzer caused some questions in the SO_2 removal results which were reported.

3.5.2 Scrubber Equipment

Some of the scrubber equipment problems consisted of downcomer line erosion, failure of the upper bed supports in one scrubber, nozzle plugging, nozzle erosion, scrubber feed ladder vane erosion, and erosion of the scrubber walls at the gas entry point.

A 1/16 inch loss of material from the fiberglass reinforced downcomer lines was noticed in August, 1973. This was due to abrasive action of the glass marbles on the drain pots. To remedy the situation, 304 stailess steel covers were placed over the fiberglass pot drains.

In September, 1973, an increase in the pressure drop across one scrubber caused a shut down. Upon inspection, it was found that the support for the upper marble bed had collapsed. All of the marbles and the upper bed support were on the lower bed. The support was welded back in place with extra care taken to protect the welds from corrosion. No evidence of support failure has been found in either scrubber since that time. This support failure occurred simultaneously with the sulfite scaling problems. It is possible that the increased pressure drop created enough stress to weaken the welds holding the supports in place.

It was also noted in September, 1973, that several spraywater nozzles were being eroded. Subsequent to this a nozzle was found whose nipple was eroded completely through. Holes were also eroded in the spraywater header. These original nozzles were made of 316 stainless steel. All spray nozzles were replaced with plastic nozzles lined with hard rubber (Nordel) linings to reduce their susceptibility to erosion.

In early 1974, it was discovered that erosion of the ladder vanes was continuing. The Ceilcote protective coating had eroded away exposing the vanes to further attack. It was reasoned that the ladder vane spray washes were causing at least part of the erosion. The frequency of wash and the water pressure used were significantly reduced. The tubing in the ladder vanes was replaced with 316 stainless steel and the Ceilcote coating was also replaced. No further erosion has been noted, but the time of operation since early 1974 has not been significant.

At the time that the ladder vane erosion was noticed, erosion was noted on the scrubber walls around the flue gas entry point. A gas inlet steam blower had been installed with the original equipment to prevent ash build up at the scrubber inlet. This is a high pressure steam blower, and several times condensed water in this line had been forced at high pressure into the scrubber walls causing the erosion. Since the electrostatic precipitator removes 95% of the ash, no ash build up is anticipated. The gas inlet steam blower has not been operated since the erosion was noted, and no ash build up has occurred.

In July, 1974, some spraywater nozzles became plugged. The pluggage was due to rust from the carbon steel feed lines. The feed lines are all carbon steel except for the portion inside the scrubber. The nozzles were cleaned, but rust plugging could be a continuing problem as the unit continues to operate.

3.5.3 Strainers and Control Valves

Early in the system's operating history, fouling of the spraywater strainers and additive control valve was a major problem. The feed slurry flow was limited and pH control was not very reliable. The source of the problem was in the lime additive. Originally, the lime slurry was centrifuged prior to its delivery to Paddy's Run No. 6. After this practice was eliminated, the level of impurities in the lime feed was greatly increased. The strainers had to be cleaned every two hours to remove a stringy material which restricted flow. Large amounts of calcium carbonate and ferrosilicon were also being introduced with the lime. A Reitz disintegrator was added on top of the additive slurry tank to grind the large calcium carbonate particles as well as reduce the size of other impurities. With the disintegrator in service, the time between backwashing the filters was increased to over 24 hours, and additive control valve plugging problems were greatly reduced.

3.5.4 Pump Service

Initially, all the pumps in the system worked well except the spraywater feed pumps. The impeller clearance in these pumps was found to be too small, which had resulted in damage to the impellers. Once clearance was increased, the pumps worked adequately.

In September, 1973, after five months of intermittent operation (approximately 1,000 hours of system operation) wear was noticed on the scrubber drain pumps. The original carbon steel throat bushing showed wear and corrosion. A 304 stainless steel replacement was fabricated and installed.

The spraywater pumps experienced two major problems. The first was noted in November, 1973. The gland seal water supply was not sufficient to lubricate two spraywater pumps. Two pumps are required when both scrubbers are operating. To correct the problem, a 5.08 cm (2 inch) gland water line was installed to replace the original 2.54 cm (1 inch) line.

The second spraywater pump problem was diagnosed after the extended run in early 1974. The pumps had lost capacity due to erosion after approximately 2,000 hours of operation. The original carbon steel pump internals were replaced with internals made of nickel alloy. Not enough operating time has passed since that time to assess the reliability of the new pump internals.

4.0 ASSESSMENT OF PREVIOUS DATA

A detailed assessment of data gathered during operation at Paddy's Run No. 6 lime scrubber is made in this section. Since the scrubbing system has been in operation for a total of about six months, it was not feasible to completely analyze all of the available data in the time allotted to this program. For this reason it was decided that it was more meaningful to attempt a complete analysis of a selected period of operation, rather than to review the entire test program in cursory fashion. The operation period which was selected for this analysis was one which was considered to be characteristic of typical LG&E system performance.

This section contains a brief discussion of the selected operating period analyzed. The results of chemical equilibrium calculations, which were used to evaluate the accuracy of reported chemical analyses, are presented along with the important process rate calculations. These results are compared to previous results from pilot plants. A preliminary analysis of the draft tube chemistry and its importance to the overall system is also made. Finally, conclusions and recommendations are presented.

4.1 Description of Data

The Paddy's Run No. 6 lime scrubbing system has been operated for a total of six months since 1973. However, for the majority of this time, complete sets of operating and analytical data are not available. The information required to perform a meaningful analysis of an SO_2 removal process includes liquid and gas phase chemical species concentrations and temperatures. The chemical species of importance for purposes of characterizing lime system scrubbing liquors are hydrogen (pH), calcium, magnesium, sodium, chloride, carbonate, sulfite, and sulfate. To perform lime dissolution and calcium sulfate and sulfite precipitation rate calculations, calcium, sulfite, and sulfate concentrations must be specified as well as the flow rates for all the streams in the scrubbing loop.

The most extensive data which were gathered during the entire test period were obtained for the time period November 28-30, 1973. November 30 was also the termination of the 35 day continuous run, which is the longest period of continuous operation of the scrubbing unit to date. A somewhat less comprehensive set of analytical data are available for the month of November. These data were sufficient to permit precipitation and dissolution rates to be calculated but were not sufficient to allow the calculation of chemical equilibria. Since the unit had been operating continuously for some time and since complete chemical analyses were available, the period from November 23 through November 30, 1973, was chosen for detailed engineering analysis. The chemical analysis data gathered during this week are listed in Tables 4-1 and 4-2.

4.2 Chemical Equilibrium Calculations

One objective of this subtask was to evaluate the completeness and accuracy of existing data. The period from November 28 through November 30, 1973, was found to have sufficient data to permit a calculation of chemical equilibria. (Sodium and chloride analyses were not available for these days so other operating periods were used to estimate these concentrations.) For the most part, data for other operating periods were not found to be complete enough to allow detailed chemical equilibria calculations to be made.

TABLE 4-1

LIQUID PHASE CHEMICAL ANALYSES

(November 23 - November 30, 1973)

			Liquid Composition (ppm)			
	pH	Temp. (°C)	Ca^{++}	Mg^{++}	$SO_3^{=}$	$SO_4^{=}$
November 23						
Upper 'B' Downcomer	-	-	552	95	721	495
Lower 'B' Downcomer	-	-	604	102	980	504
Hold Tank Effluent	-	-	428	75	63	604
November 24						
Upper 'B' Downcomer	-	-	528	104	823	473
Lower 'B' Downcomer	-	-	588	102	1,023	573
Hold Tank Effluent	-	-	408	87	121	555
November 25						
Upper 'B' Downcomer	-	-	404	99	726	689
Lower 'B' Downcomer	-	-	500	58	1,142	670
Hold Tank Effluent	-	-	368	58	82	882
November 26						
Lower 'A' Downcomer	-	-	732	24	1,223	433
Lower 'B' Downcomer	-	-	656	44	1,039	434
Hold Tank Effluent	-	-	432	44	101	419
November 27						
Upper 'B' Downcomer	5.2	-	528	29	600	688
Lower 'B' Downcomer	5.3	-	608	29	941	740
Hold Tank Effluent	8.0	-	448	29	63	584
November 28						
Upper 'B' Downcomer	5.3	48	480	90	606	442
Lower 'B' Downcomer	5.0	47	558	96	845	514
Upper 'A' Downcomer	5.3	46	467	90	454	556
Lower 'A' Downcomer	4.9	47	465	88	986	-
'A' Scrubber Bottoms	4.9	45	583	108	892	691
'B' Scrubber Bottoms	4.9	45	588	109	872	692
Hold Tank Effluent	8.0	45	396	78	82	467
Filtrate	4.9	35	533	109	540	790
Clarifier Underflow	-	-	485	113	610	318
Clarifier Overflow	-	-	594	108	790	10

TABLE 4-1 LIQUID PHASE CHEMICAL ANALYSES (Cont.)

			Liquid Composition (ppm)			
	pH	Temp. (°C)	Ca^{++}	Mg^{++}	$SO_3^{=}$	$SO_4^{=}$
November 29						
Upper 'B' Downcomer	5.3	49	533	109	704	602
Lower 'B' Downcomer	5.1	49	629	114	1,030	637
Upper 'A' Downcomer	5.4	48	521	111	622	618
Lower 'A' Downcomer	5.0	48	679	111	1,255	605
'A' Scrubber Bottoms	5.0	44	641	129	1,033	731
'B' Scrubber Bottoms	5.0	44	630	128	997	722
Hold Tank Effluent	8.0	46	420	94	66	582
Filtrate	-	-	502	133	426	838
Clarifier Underflow	-	-	664	133	-	-
Clarifier Overflow	-	-	605	129	810	813
November 30						
Upper 'B' Downcomer	5.2	49	532	118	692	618
Lower 'B' Downcomer	4.9	41	629	123	1,064	576
Upper 'A' Downcomer	5.3	48	520	119	661	602
Lower 'A' Downcomer	4.9	48	661	123	1,181	669
'A' Scrubber Bottoms	4.9	44	632	135	1,119	636
'B' Scrubber Bottoms	4.9	45	614	136	986	722
Hold Tank Effluent	7.7	47	418	102	78	600
Filtrate	-	-	526	147	548	831
Clarifier Underflow	5.3	43	937	147	-	-
Clarifier Overflow	4.8	32	619	139	837	857

TABLE 4-2

SOLID PHASE CHEMICAL ANALYSES

(November 28 - November 30, 1973)

	Solids Content (% of solids total)				
	Ca	Mg	SO_3	SO_4	CO_2
November 28					
Upper 'B' Downcomer	28.1	0.1	46.2	3.2	-
Lower 'B' Downcomer	28.1	0.1	45.5	4.4	-
Upper 'A' Downcomer	28.1	0.1	46.2	3.9	3.6
Lower 'A' Downcomer	27.9	0.2	45.9	3.4	-
'A' Scrubber Bottoms	27.4	0.1	45.5	3.4	-
'B' Scrubber Bottoms	27.1	0.2	44.5	3.1	-
Hold Tank Effluent	27.9	0.2	45.5	2.8	3.5
Filter Cake	27.2	0.2	45.5	3.4	-
Clarifier Underflow	27.2	0.2	45.5	2.9	-
November 29					
Upper 'B' Downcomer	27.0	0.2	45.1	3.9	-
Lower 'B' Downcomer	26.9	0.1	43.8	3.2	4.0
Upper 'A' Downcomer	27.0	0.2	45.0	3.5	-
Lower 'B' Downcomer	27.0	0.2	44.7	3.5	-
'A' Scrubber Bottoms	26.8	0.1	45.3	3.1	-
'B' Scrubber Bottoms	26.6	0.2	45.1	3.8	-
Hold Tank Effluent	26.8	0.2	44.9	2.8	3.3
Filter Cake	26.6	0.1	44.8	3.6	-
Clarifier Underflow	26.4	0.2	45.0	3.6	-
November 30					
Upper 'B' Downcomer	26.9	0.1	45.0	3.7	-
Lower 'B' Downcomer	27.1	0.1	44.4	4.4	3.7
Upper 'A' Downcomer	26.8	0.1	44.7	3.9	-
Lower 'A' Downcomer	26.9	0.1	44.8	3.9	-
'A' Scrubber Bottoms	26.4	0.1	44.7	3.4	-
'B' Scrubber Bottoms	26.5	0.1	44.9	3.4	-
Hold Tank Effluent	26.9	0.1	44.9	3.2	2.7
Filter Cake	26.4	0.1	44.9	3.2	-
Clarifier Underflow	26.4	0.1	44.9	3.3	-

The Radian chemical equilibrium program was utilized in an effort to check the accuracy of reported chemical analyses. The results of these equilibria runs are listed in Table 4-3. Residual electroneutrality is the parameter which can be used as a measure of the consistency of the data. The electroneutrality imbalance is reported as the ratio of the charge imbalance to the total equivalents of charged species present in a given sample.

It has been Radian's experience that a three to five percent charge imbalance is an indication of reasonably accurate analytical data. A ten percent electroneutrality error is probably the maximum deviation allowable. Using this 10% figure as a guideline, the majority of the LG&E chemical data is judged to be adequate. However, roughly one fourth of the data reported by Combustion Engineering for the time period analyzed (28-30 November 1973) showed residual electroneutralities in excess of 10%. For all three days, the spraywater feed stream charge imbalances were in excess of 15%. The other streams for which large imbalances were noted are (1) upper 'B' downcomer on 11/28/73 (14.8%), (2) upper 'A' downcomer on 11/28/73 (11.5%), and (3) lower 'B' downcomer on 11/30/73 (11.0%).

All of the residual electroneutralities calculated were positive in value, which indicates an excess of positive charges. One possible error is that no carbonate or nitrate ion analyses were performed. As a result the carbonate and nitrate inputs to the Radian equilibrium program were specified to be zero. These would account for some of the charge imbalance errors which are observed in the data. Sorption of relatively small quantities of CO_2 and NO_x in the flue gas could result in fairly significant carbonate and nitrate ion concentrations in the circulating liquor. Without independent tests being performed, it is impossible to determine the exact cause of the electroneutrality imbalances.

TABLE 4-3
LIQUID PHASE ANALYSES FOR PADDY'S RUN NO. 6 LIME SCRUBBER

Process Streams	Temp. (^oC)	pH	Concentrations, moles/liter: Ca	Mg[1]	Na_2[2]	Cl[2]	SO_2	SO_3	charge imbalance (%)	Relative Saturations: $CASO_3 \cdot \frac{1}{2}H_2O$	$CaSO_4 \cdot 2H_2O$
November 28, 1973											
Upper 'A' Downcomer	46	5.3	.0117	.0046	.0005	.01	.0057	.0058	11.5	2.00	.283
'A' Scrubber Bottoms	45	4.9	.0146	.0046	.0005	.01	.0110	.0072	5.6	1.95	.383
Upper 'B' Downcomer	48	5.3	.012	.0046	.0005	.01	.0076	.0058	14.8	2.68	.228
Lower 'B' Downcomer	47	5.0	.0135	.0046	.0005	.01	.0106	.00535	8.9	2.20	.281
'B' Scrubber Bottoms	45	4.9	.0147	.0046	.0005	.01	.0109	.0072	6.1	1.95	.385
Hold Tank Effluent	45	8.0	.0099	.0046	.0005	.01	.0010	.0045	20.3	2.78	.222
Filtrate	35	4.9	.0133	.0046	.0005	.01	.0068	.0082	5.8	1.31	.433
November 29, 1973											
Upper 'A' Downcomer	48	5.4	.013	.0047	.0005	.01	.0078	.0064	8.8	3.37	.318
Lower 'A' Downcomer	48	5.0	.017	.0047	.0005	.01	.0160	.0063	6.7	3.75	.354
'A' Scrubber Bottoms	44	5.0	.016	.0047	.0005	.01	.0130	.0076	5.0	3.05	.417
Upper 'B' Downcomer	49	5.3	.0133	.0047	.0005	.01	.0088	.0063	7.3	3.08	.313
Lower 'B' Downcomer	49	5.1	.0157	.0047	.0005	.01	.0130	.0066	7.9	3.53	.361
'B' Scrubber Bottoms	44	5.0	.0158	.0047	.0005	.01	.0120	.0075	6.5	2.81	.411
Hold Tank Effluent	46	8.0	.0105	.0047	.0005	.01	.0008	.0061	18.1	2.22	.281
November 30, 1973											
Upper 'A' Downcomer	48	5.3	.013	.005	.0005	.01	.0083	.0063	9.4	2.96	.311
Lower 'A' Downcomer	48	4.9	.017	.005	.0005	.01	.0150	.0069	7.7	2.84	.384
'A' Scrubber Bottoms	44	4.9	.016	.005	.0005	.01	.0140	.0066	7.5	2.72	.363
Upper 'B' Downcomer	49	5.2	.0133	.005	.0005	.01	.0087	.0064	8.5	2.50	.316
Lower 'B' Downcomer	41	4.9	.016	.005	.0005	.01	.0130	.006	11.0	2.70	.338
'B' Scrubber Bottoms	45	4.9	.015	.005	.0005	.01	.0120	.0075	5.3	2.14	.396
Hold Tank Effluent	47	7.7	.010	.005	.0005	.01	.0010	.0062	16.1	2.59	.270
Clarifier Overflow	32	4.8	.015	.005	.0005	.01	.0100	.0089	4.3	1.76	.494

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[1] Magnesium concentrations were averaged for all streams since it should remain constant for the system.

[2] These analyses not available for this time period. Data from another period were used to estimate these concentrations.

Calculated calcium sulfate and calcium sulfite relative saturations for the various process streams are also listed in Table 4-3. Relative saturation is a measure of the extent to which the activity product of a precipitating species exceeds the equilibrium solubility of the solid formed. The relative saturation of calcium sulfite hemihydrate for example is

$$\text{R.S.} = \frac{a_{\text{Ca}^{++}}\, a_{\text{SO}_3^{=}}\, a_{\text{H}_2\text{O}}^{\frac{1}{2}}}{K_{\text{SP, CaSO}_3\cdot\frac{1}{2}\text{H}_2\text{O}}} \qquad (4\text{-}1)$$

The relative saturation figures presented in Table 4-3 indicate that all of the streams considered were unsaturated with respect to calcium sulfate dihydrate. These saturations ranged from .2 to .5 of the equilibrium solubility. Operation at these calcium sulfate levels eliminates the chances of sulfate scaling. These results can be explained by the low chloride concentrations and low oxidation levels observed in the LG&E system. Pilot scale experiments conducted at low chloride concentrations and low oxidation levels have produced results which support the existence of the subsaturated conditions occurring at LG&E (BO-147). The subsaturation is made possible by the purging of calcium sulfate as a solid solution with calcium sulfite.

The calcium sulfite relative saturations were observed to fluctuate rather significantly from day to day. On November 28, 1974, calculated calcium sulfite relative saturations on the marble beds ranged from 2.0 to 2.8. On November 29 these same relative saturations ranged from 3.0 to 3.8 on the same beds. These variations would normally be the result of some change in a critical operating paramter such as the SO_2 level in the flue gas. However, on the days under consideration no such large fluctuations occurred. Furthermore, no pH fluctuations were noted throughout the system. Sulfite relative saturations above 4 approach possible (but not necessarily definite) scaling levels (OT-023).

4.3 Process Rate Calculations

Precipitation and dissolution rate calculations were employed in analyzing the performance of the LG&E lime based flue gas desulfurization system. Evaluation of these rates for the significant process vessels provided insight into the process chemistry of the system. As discussed in Section 4.2, the chemical equilibria calculations and electroneutrality balances which were made indicated that liquid phase analytical data were, for the most part, complete and accurate. By combining operating information with the chemical data, important rates were calculated for the scrubbers and hold tank.

In addition to the SO_2 pickup rate, the four key reactions whose rates are instrumental in characterizing a lime scrubbing system are:

sulfur dioxide oxidation*

$$SO_2(aq) + \tfrac{1}{2}O_2 \xrightarrow{R_1} SO_3(aq) \qquad (4\text{-}2)$$

calcium sulfite precipitation

$$Ca^{++} + SO_3^{=} + \tfrac{1}{2}H_2O \xrightarrow{R_2} CaSO_3 \cdot \tfrac{1}{2}H_2O(s) \qquad (4\text{-}3)$$

calcium sulfate precipitation**

$$Ca^{++} + SO_4^{=} + 2H_2O \xrightarrow{R_3} CaSO_4 \cdot 2H_2O(s) \qquad (4\text{-}4)$$

---

*Reactive form of S(IV) is unknown. HSO_3^- or $SO_3^=$ for example, may actually be the component(s) which is oxidized.

**Calcium sulfate may actually precipitate as a solid solution whose exact hydration form is unknown.

calcium hydroxide or calcium carbonate dissolution*

$$Ca(OH)_2(s) \xrightarrow{R_4} Ca^{++} + 2OH^- \qquad (4\text{-}5)$$

$$CaCO_3(s) \rightarrow Ca^{++} + CO_3^{=}$$

The SO_2 which is absorbed from the flue gas is removed by oxidation (R_1) and by precipitation (R_2). Sulfur species formed by oxidation are removed by precipitation (R_3). (Small amounts of these species also exit the system as soluble components in the liquor associated with the solid filter cake.) The lime is added to the scrubbing liquor by dissolution (R_4).

In principle, these reaction rates can be calculated utilizing either solid or liquid species material balances. However, because of the 10% solids in the scrubbing slurry, changes in the solids due to precipitation cannot be determined to any acceptable degree of accuracy. Table 4-4 demonstrates this by showing the sulfur content of the various streams around the scrubbers. It is evident that the quantity of sulfur in the solid phase of the feed greatly exceeds the quantity of sulfur which precipitates from the liquid. The potential error in the solids analysis is at least as great as the quantity of sulfur which precipitates. Also, the liquid analyses are much more complete than are the solids analyses for the Louisville system. For these reasons liquid species balances were used in performing the rate calculations.

In the following subsections the assumptions required to perform these rate calculations, the computational procedures, and the possible sources of errors are discussed. A summary of results is also presented.

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*Since liquid carbonate concentration measurements were not made, it is not possible to differentiate between these reactions.

TABLE 4-4

TOTAL SULFUR MATERIAL BALANCE AROUND THE SCRUBBERS

S balance		Stream Flow Rate	S in Liquid (m moles/l)	Total Solids Conc.[1] (g/l)	S in Solids (m moles/g)	S in Gas (ppm)	Total S in Liquid (g moles/min)	Total S in Gas (g moles/min)	Total S in Solids (g moles/min)	Total S (g moles/min)
Inlet	Flue gas inlet	266,400 g moles/min	-	108	-	1,875	-	500	-	500
	Scrubber feed	45,360 l/min	7.23	108	5.94	-	330	-	29,100	29,430
Outlet	Flue gas outlet	279,800 g moles/min	-	108	-	275[2]	-	80	-	80
	Upper 'A' Downcomer	7,560 l/min	14.5	108	6.0	-	110	-	4,900	5,010
	Lower 'A' Downcomer	11,340 l/min	21.8	108	6.01	-	250	-	7,360	7,720
	'A' Scrubber Bottoms	3,780 l/min	20.6	108	5.94	-	80	-	2,420	2,500
	Upper 'B' Downcomer	7,560 l/min	14.9	108	6.01	-	110	-	4,910	5,020
	Lower 'B' Downcomer	11,340 l/min	19.3	108	6.0	-	220	-	7,350	7,570
	'B' Scrubber Bottoms	3,780 l/min	19.8	108	5.96	-	80	-	2,430	2,510

In = 29,925 g moles/min

Out= 30,400 g moles/min

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1. Average solids concentration for November 30, 1973.
2. The outlet SO_2 concentration was averaged over both scrubbers.

4.3.1 Assumptions

Certain assumptions were necessary to calculate the precipitation and dissolution rates. These are presented below with justifications discussed where necessary.

4.3.1.1 System Oxidation

From equations 4-2 through 4-5 it is evident that the oxidation rate (R_1) is required to calculate the remaining rates using liquid species balances. Observations at Combustion Engineering's prototype facility showed that oxidation in the hold tank was negligible (OT-023). The percent oxidation occurring in any hold tank is proportional to the surface area to volume ratio for the tank.

$$\%\ \text{oxidation}\ \alpha\ \frac{\text{surface area}}{\text{volume}} = \frac{\pi D^2/4}{H\pi D^2/4} \qquad (4\text{-}6)$$

where: D is the diameter
H is the height

This reduces to:

$$\%\ \text{oxidation}\ \alpha\ \frac{1}{H} \qquad (4\text{-}7)$$

The hold tank at Louisville is 832,000 liters compared to a 22,700 liter hold tank at Windsor. If the two tanks are assumed to have proportional dimensions, then the Louisville tank would be expected to have a lower percent oxidation since its surface to volume ratio is less. Therefore, an assumption that no significant oxidation occurred in the hold tank at LG&E is probably reasonable.

4.3.1.2 SO_2 Removal Efficiency

Process rate calculations were performed for the individual marble beds. Since no gas samples were taken between the beds, an SO_2 removal efficiency had to be assumed for each bed. Data from Combustion Engineering's Kreisinger Development Laboratory shows that SO_2 removal efficiency is primarily a function of marble bed pH (OT-023). Data for the LG & E system indicates that the pH did not vary significantly between the two beds. Therefore, an assumption of equal removal efficiencies for the two beds in the scrubber was made.

4.3.1.3 Downcomer Flow Rates

Flows from the marble beds through the downcomers to the hold tank are controlled by gravity. Since these large lines do not operate in a liquid full manner, no measurements of downcomer flow rates are available. It was assumed that half of the total slurry flow rates exits through the lower bed downcomer, one-third of the total exits through the upper bed downcomer, and one-sixth of the flow falls into the scrubber tank. The bases for this assumption are Combustion Engineering's design criteria which are supported by Kreisinger lab data (BU-157).

4.3.1.4 Bed Reject Composition and Flow Rate

When individual bed rates and performance are analyzed, bed reject or "weep" compositions and flow rates are required. An assumption of equal "weep" flow from each bed was made. The flow was assumed to be 1/6 of the total flow to the scrubber. This follows directly from the assumed downcomer flow rates. The bed reject composition was assumed to be identical to that for downcomer liquor exiting the same bed.

4.3.1.5 Scrubber Bottoms Compositions

The scrubber bottoms were not sampled during normal sampling periods. The scrubber bottoms composition was assumed to be the same as the lower bed downcomer's composition when analytical data were unavailable. This was judged to be a reasonable assumption after evaluating the chemical analysis data for those days when both streams where analyzed (Table 4-3).

4.3.1.6 Clarifier Overflow and Lime Additive Liquid Compositions

Since no chemical analyses were available for these streams, the clarifier overflow and lime additive streams were neglected in the liquid balances. These streams are small compared to the other streams entering the hold tank so the error due to this assumption is negligible.

4.3.2 Computational Procedure

All of the oxidation in the system has been assumed to occur in the scrubber (R_1-hold tank = 0). Combining this with the assumption that there is negligible SO_3 in the flue gas yields, at steady state, Equation 4-8.

$$\text{fraction } SO_2 \text{ oxidized} = \left(\frac{SO_2 \text{ oxidized in the system}}{SO_2 \text{ absorbed}}\right) = \left(\frac{SO_3(s)}{SO_2(s) + SO_3(s)}\right)_{\text{leaving the system}} \tag{4-8}$$

The oxidation rate becomes

$$R_1 = \left(SO_2(g)_{in} - SO_2(g)_{out}\right)\left(\frac{SO_3(s)}{SO_2(s) + SO_3(s)}\right) \tag{4-9}$$

Using this expression for oxidation rate the calcium sulfate and sulfite precipitation rates and calcium hydroxide/ calcium carbonate dissolution rate can be calculated. Equations 4-10 through 4-12 are the appropriate molar rate equations which were derived from the liquid species balances. These equations hold for any process vessel.

$$\text{calcium sulfite precipitation} = R_2 = SO_2(aq + g)_{in} - SO_2(aq + g)_{out} - R_1 \qquad (4\text{-}10)$$

$$\text{calcium sulfate precipitation} = R_3 = SO_3(aq)_{in} - SO_3(aq)_{out} + R_1 \qquad (4\text{-}11)$$

$$\text{calcium hydroxide and/or calcium carbonate dissolution} = R_4 = Ca(aq)_{out} - Ca(aq)_{in} + R_2 + R_3 \qquad (4\text{-}12)$$

4.3.3 Summary of Results

Typical molar oxidation rates were about five percent during November 28-30, 1973 (Table 4-2). Table 4-5 presents a rate summary for November 23-30, 1973. Figure 4-1 gives a pictorial representation for the rates calculated for November 30, 1973. Appendix I shows a sample of the detailed rate results which were calculated by treating the scrubber as a single unit. Table 4-6 presents the rates for the individual marble beds.

The reasonabless of these calculated rates may be checked using several criteria. Two of these are (at steady state):

TABLE 4-5

LG & E RATE CALCULATIONS SUMMARY

(Based on liquid species balances)

	Rates (g moles/min)						
Scrubber(s)	11/23	11/24	11/25	11/27	11/28	11/29	11/30
SO_2 Oxidation (assumed)	11.7	11.7	12.0	9.9	19.3	19.4	23.6
$CaSO_3$ Precipitation	-53.1	-54.4	-73.6	-62.0	-22.9	-130.0	-102.4
$CaSO_4$ Precipitation	35.2	22.5	60.5	-22.4	-30.5	-5.7	10.6
$Ca(OH)_2$ and/or $CaCO_3$ Dissolution	72.3	58.9	43.6	-8.8	77.3	78.0	113.8
SO_2 Absorption	194.7	194.7	200.0	165.1	392.7	397.7	422.6
Hold Tank							
$CaSO_3$ Precipitation	236.1	237.4	261.6	217.2	396.5	508.2	501.5
$CaSO_4$ Precipitation	-23.5	-10.8	-48.5	32.3	49.5	25.1	13.0
$Ca(OH)_2$ and/or $CaCO_3$ Dissolution	122.4	135.8	156.4	173.9	315.3	319.7	308.8
System Balance							
Total Sulfur Precipitation	194.7	194.7	200.0	165.1	392.6	397.5	422.7
$Ca(OH)_2$ and/or $CaCO_3$ Dissolution	194.7	194.7	200.0	165.1	392.6	397.7	422.6
SO_2 Absorption	194.7	194.7	200.0	165.1	392.7	397.7	422.6

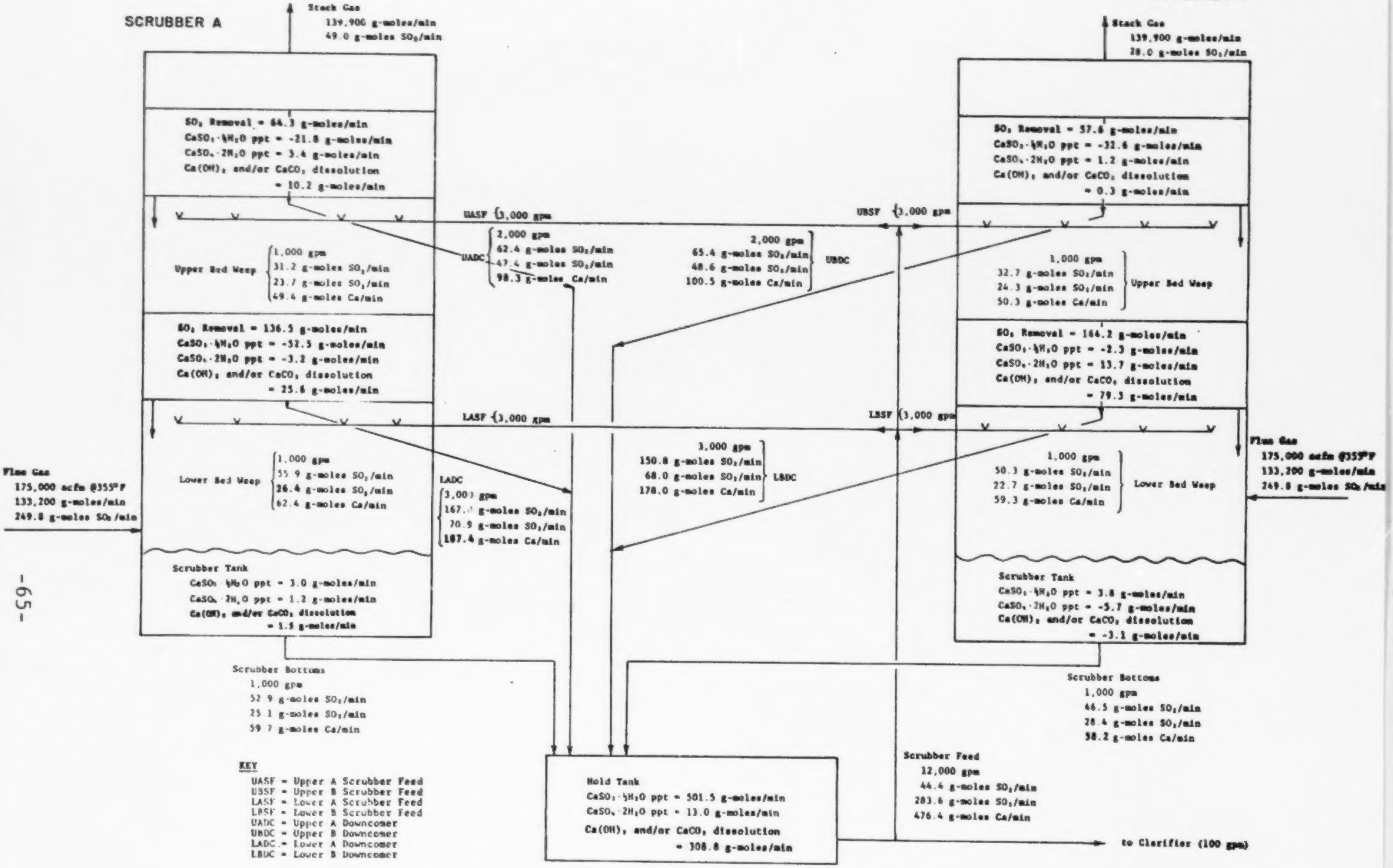
SCRUBBER A
Stack Gas
139,900 g-moles/min
49.0 g-moles SO_2/min
SO_2 Removal = 64.3 g-moles/min
$CaSO_3 \cdot \frac{1}{2}H_2O$ ppt = -21.8 g-moles/min
$CaSO_4 \cdot 2H_2O$ ppt = 3.4 g-moles/min
$Ca(OH)_2$ and/or $CaCO_3$ dissolution = 10.2 g-moles/min
Upper Bed Weep
1,000 gpm
31.2 g-moles SO_3/min
23.7 g-moles SO_4/min
49.4 g-moles Ca/min
SO_2 Removal = 136.5 g-moles/min
$CaSO_3 \cdot \frac{1}{2}H_2O$ ppt = -32.5 g-moles/min
$CaSO_4 \cdot 2H_2O$ ppt = -3.2 g-moles/min
$Ca(OH)_2$ and/or $CaCO_3$ dissolution = 25.6 g-moles/min
Lower Bed Weep
1,000 gpm
55.9 g-moles SO_3/min
26.4 g-moles SO_4/min
62.4 g-moles Ca/min
Flue Gas
175,000 acfm @355°F
133,200 g-moles/min
249.8 g-moles SO_2/min
Scrubber Tank
$CaSO_3 \cdot \frac{1}{2}H_2O$ ppt = 3.0 g-moles/min
$CaSO_4 \cdot 2H_2O$ ppt = 1.2 g-moles/min
$Ca(OH)_2$ and/or $CaCO_3$ dissolution = 1.5 g-moles/min
Scrubber Bottoms
1,000 gpm
52.9 g-moles SO_3/min
25.1 g-moles SO_4/min
59.7 g-moles Ca/min
UASF {3,000 gpm
UADC
2,000 gpm
62.4 g-moles SO_3/min
47.4 g-moles SO_4/min
98.3 g-moles Ca/min
LASF {3,000 gpm
LADC
3,000 gpm
167.0 g-moles SO_3/min
70.9 g-moles SO_4/min
187.4 g-moles Ca/min
UBSF {3,000 gpm
UBDC
2,000 gpm
65.4 g-moles SO_3/min
48.6 g-moles SO_4/min
100.5 g-moles Ca/min
LBSF {3,000 gpm
LBDC
3,000 gpm
150.8 g-moles SO_3/min
68.0 g-moles SO_4/min
178.0 g-moles Ca/min
SCRUBBER B
Stack Gas
139,900 g-moles/min
28.0 g-moles SO_2/min
SO_2 Removal = 57.6 g-moles/min
$CaSO_3 \cdot \frac{1}{2}H_2O$ ppt = -32.6 g-moles/min
$CaSO_4 \cdot 2H_2O$ ppt = 1.2 g-moles/min
$Ca(OH)_2$ and/or $CaCO_3$ dissolution = 0.3 g-moles/min
Upper Bed Weep
1,000 gpm
32.7 g-moles SO_3/min
24.3 g-moles SO_4/min
50.3 g-moles Ca/min
SO_2 Removal = 164.2 g-moles/min
$CaSO_3 \cdot \frac{1}{2}H_2O$ ppt = -2.3 g-moles/min
$CaSO_4 \cdot 2H_2O$ ppt = 13.7 g-moles/min
$Ca(OH)_2$ and/or $CaCO_3$ dissolution = 79.3 g-moles/min
Lower Bed Weep
1,000 gpm
50.3 g-moles SO_3/min
22.7 g-moles SO_4/min
59.3 g-moles Ca/min
Flue Gas
175,000 acfm @355°F
133,200 g-moles/min
249.8 g-moles SO_2/min
Scrubber Tank
$CaSO_3 \cdot \frac{1}{2}H_2O$ ppt = 3.8 g-moles/min
$CaSO_4 \cdot 2H_2O$ ppt = -5.7 g-moles/min
$Ca(OH)_2$ and/or $CaCO_3$ dissolution = -3.1 g-moles/min
Scrubber Bottoms
1,000 gpm
46.5 g-moles SO_3/min
28.4 g-moles SO_4/min
58.2 g-moles Ca/min
Hold Tank
$CaSO_3 \cdot \frac{1}{2}H_2O$ ppt = 501.5 g-moles/min
$CaSO_4 \cdot 2H_2O$ ppt = 13.0 g-moles/min
$Ca(OH)_2$ and/or $CaCO_3$ dissolution = 308.8 g-moles/min
Scrubber Feed
12,000 gpm
44.4 g-moles SO_3/min
283.6 g-moles SO_4/min
476.4 g-moles Ca/min
to Clarifier (100 gpm)
KEY
UASF = Upper A Scrubber Feed
UBSF = Upper B Scrubber Feed
LASF = Lower A Scrubber Feed
LBSF = Lower B Scrubber Feed
UADC = Upper A Downcomer
UBDC = Upper B Downcomer
LADC = Lower A Downcomer
LBDC = Lower B Downcomer

FIGURE 4-1. RATE CALCULATIONS USING LIQUID SPECIES BALANCE

(Paddy's Run No. 6 Lime Scrubbing System - November 30, 1973)

TABLE 4-6

INDIVIDUAL MARBLE BED RATE SUMMARY

(All values in g moles/min)

	$CaSO_3$ Precipitation		$CaSO_4$ Precipitation		$Ca(OH)_2$ and/or $CaCO_3$ Dissolution		SO_2 Absorption	
	11/29	11/30	11/29	11/30	11/29	11/30	11/29	11/30
Scrubber 'B'								
Upper Bed	-35.9	-32.6	0.4	1.2	-3.8	0.3	57.3	57.6
Lower Bed	-8.1	-2.3	-0.6	13.7	59.4	79.3	151.7	164.2
Scrubber Tank	1.5	3.8	-3.3	-5.7	-1.3	-3.1	-	-
TOTAL	-42.5	-31.1	-3.5	9.2	54.3	76.5	209.0	221.8
Scrubber 'A'								
Upper Bed	-19.9	-21.8	-1.2	3.4	7.2	10.2	62.2	64.3
Lower Bed	-78.2	-52.5	4.0	-3.2	14.7	25.6	126.5	136.5
Scrubber Tank	10.6	3.0	-5.0	1.2	1.8	1.5	-	-
TOTAL	-87.5	-71.3	-2.2	1.4	23.7	37.3	188.7	200.8

- The sum of sulfite and sulfate precipitation rates in the scrubber and hold tank should not exceed the system SO_2 removal rate.

- The calcium hydroxide dissolution rate should equal or slightly exceed the sum of the sulfite and sulfate precipitation rates for the system. (The difference is primarily the calcium carbonate that is precipated when CO_2 is absorbed in the system.)

Using these criteria overall system reaction rates appear to be reasonably consistent. However, in reviewing the scrubber and hold tank rates some questionable numbers are noted, particularly on November 23, 24, and 25. On these days, significant sulfite dissolution rates and relatively large sulfate precipitation rates were calculated for the scrubber. Since all of streams are subsaturated with respect to gypsum, this does not appear to be reasonable. These differences in the signs of calculated precipitation rates seem to indicate an analytical error.

Another indication of probable analytical error is the apparent dissolution of calcium sulfate which occurred in the hold tank on these days. Although the coprecipitation phenomenon is not completely understood, it is more likely that calcium sulfate forms a solid solution with calcium sulfite when there is a net precipitation of sulfite. The probability, therefore, is that the majority of the sulfate precipitates in the hold tank with the sulfite.

The assumption that all of the oxidation occurred in the scrubber may have contributed to the calcium sulfate-calcium sulfite precipitation inconsistencies. It is not possible to verify this assumption for the LG&E system based on existing data. If any oxidation actually occurred in the hold tank rather than the scrubber, the apparent calcium sulfate precipitation in the scrubber would be reduced and the $CaSO_4$ precipitation in the hold tank would be increased. However, the magnitude of the system oxidation rate is not sufficiently large to explain all of the sulfite-sulfate precipitation anomalies.

It is also noted (Table 4-5) that calcium sulfite dissolution is calculated in the scrubbers for all days considered. Yet, Table 4-3 shows that all of the scrubber effluent streams were supersaturated for November 28-30. These results are not consistent. There is no driving force for calcium sulfite dissolution above a relative saturation of one. Therefore, it is impossible for solid calcium sulfite to dissolve in a stream which is supersaturated with respect to calcium sulfite. An analytical error is probably responsible for this inconsistency.

Another inconsistency in the data is that under similar operating conditions the calculated calcium sulfite dissolutions were different for the two scrubbers (see Table 4-6). For example, on November 30, 1973, the total solid $CaSO_3$ which apparently dissolved in Scrubber B was 31.3 gmoles/min while the apparent $CaSO_3$ dissolution in Scrubber A was 71.3 gmoles/min. The rates of sulfur dioxide removal for the two scrubbers were 209 gmoles/min in Scrubber B and 188.7 in Scrubber A. It would be expected that the dissolution would be greater in Scrubber B since the SO_2 absorption is greater in Scrubber B. However, the total alkalinity provided by solids dissolution (calcium sulfite and other calcium based alkalinity) was approximately the same.

The total calculated calcium dissolved in Scrubber B was 107.6 gmoles/min and 108.6 gmoles/min in Scrubber A. This does not explain the difference in apparent sulfite dissolution for the two scrubbers, but it does show that the total alkalinity derived from solids dissolution was approximately the same.

4.3.4 Source of Possible Errors

There are several potential sources of error associated with the rate calculations and chemical equilibria runs reported above. One source mentioned previously is analytical error. Some of the data are questionable as shown by the large residual electroneutralities. These same errors are carried over into the rate calculations and are possibly magnified. It should also be mentioned that only on November 28, 29, and 30, 1973, were sufficient analytical data available to perform chemical equilibria calculations. Data taken on November 23, 24, 25, and 27 were complete enough to perform calcium sulfate and calcium sulfite rate calculations but not comprehensive enough to allow the computations of chemical equilibria. It is considered to be likely that some of these data are in error.

Another possiblity for error is sampling technique. This is particularly critical when sampling around the scrubber. Some of sample lines at Paddy's Run No. 6 exceed 50 feet in length. Since the residence time on each marble bed is only about thirty seconds, sampling as closely as possible to the marble bed is critical for good results. Significant amounts of reaction can occur in long sample lines. These reactions will yield misleading information about rates in the scrubber. Poor sampling will not be detected by chemical equilibria calculations. The analytical results can show completeness and self consistency, but still be erroneous.

A related sampling error could result from improper flushing of the sample lines. Again, long sample lines probably increase the chances of poor flushing. If the sample lines are not completely cleared of previously standing slurry, the results will not be accurate and any conclusions drawn from the analysis of such data are not meaningful.

Sample timing is also important in obtaining realistic information. The scrubber slurry feed and outlet samples should be taken simultaneously. As a rule, this was not done at the Paddy's Run No. 6 scrubbing unit. No sample times were reported for the detailed analysis performed November 28 through November 30, 1973. However, the daily sampling journals show that samples taken from various parts of the scrubber loop were normally taken from ten to fifteen minutes apart. Ten minutes is roughly 20 times the liquor residence time on a marble bed. Significant changes in bed liquor compositions can therefore occur in ten minutes. If only limited manpower is available for sampling, key parameters such as SO_2 concentration in the flue gas, should be checked closely during the sampling period. Rate calculations derived from the samples taken should be reasonably accurate if the system remains close to steady state. If steady state is not achieved, the calculations have little meaning.

4.4 Comparison of Results with Previous Data

The results of the chemical equilibria calculation show that the Paddy's Run No. 6 scrubber system operates subsaturated with respect to gypsum. This fact combined with the existence of calcium sulfate in the filter cake is evidence that sulfate-sulfite coprecipitation does occur at Paddy's Run. At Combustion Engineering's prototype facility in Windsor, calcium sulfate subsaturation was not in evidence.

One explanation for these differences in saturation levels is the amount of oxidation which occurred in each system. At the Windsor facility the oxidation was from 30 to 40%. The oxidation level at LG & E was typically less than 10%. At low chloride and magnesium levels, 20% is the maximum oxidation that can be tolerated for subsaturated operation (BO-147). Since the pH on the marble beds at the two installations were approximately the same and the sulfur to oxygen ratio was not substantially different, the cause of the low oxidation level at Louisville can only be surmised. One potential difference is the lime. Possible oxidation inhibitors or promoters could have entered the systems with the lime slurries. However, R. H. Borgwardt found no evidence of any difference in oxidation that could be attributed to differences between carbide lime and high calcium content lime (BO-146).

The rate calculations show that substantial amounts of calcium sulfite dissolve in the scrubber. (Again, it is noted that the scrubber effluent streams were calculated to be supersaturated. These results are inconsistent.) Supporting bench scale results generated at EPA indicate that calcium sulfite dissolution in the scrubber is typical of a lime system (BO-146). In fact, calcium sulfite dissolution in lime systems provides alkalinity in the same fashion as the dissolution of calcium carbonate provides alkalinity in limestone systems. Utilization of this solid phase alkalinity is a desirable feature of SO_2 scrubbing system operation.

$$CaSO_3(s) + SO_2(g) + H_2O \rightarrow Ca^{++} + 2HSO_3^-(aq) \qquad (4\text{-}6)$$

It can be seen from this equation that calcium sulfite acts as a base in the SO_2 sorption reaction.

4.5 Precipitation in the Draft Tube

One key piece of equipment which must be considered is the draft tube. The largest driving forces for dissolution and precipitation exist in the draft tube. However, no samples were taken at the outlet of the draft tube. Since no rates could be calculated and since the draft tube's function is felt to be of vital importance, the sulfite precipitation rate was estimated.

The chemical equilibrium program was used in an attempt to evaluate draft tube chemistry. Typical hold tank inlet stream chemical species compositions were put to the program and the amount of dissolved calcium hydroxide was varied. Solids precipitation was not allowed in an effort to estimate supersaturations. The results of these equilibria runs are listed in Table 4-7. As can be seen from the table, the maximum calcium sulfite relative saturation in the draft tube that can reasonably be attained is below forty. At this point, the pH is 11.0 and the magnesium hydroxide relative saturation is over 2,500.

The working volumes of the various tanks are shown in Table 4-8. Since the slurry concentration was generally around 10%, the amount of solids in the scrubbing loop (excluding the draft tube) was approximately 1.1×10^{8} grams. From Radian's experience at Pennsylvania Power and Light's pilot plant at Sunbury, a normal hold tank sulfite precipitation rate of 2×10^{-6} g moles/gram-minute of seed is estimated. Therefore, an estimate of the sulfite precipitation that could have occurred in the major tanks is 220 g moles/min.

TABLE 4-7

DRAFT TUBE SUPERSATURATIONS AT VARYING LIME ADDITIVE LOADINGS

Chemical Species Inputs (g moles/liter)

MgO	4.6×10^{-3}	SO_2	1.2×10^{-2}
Na_2O	5.0×10^{-4}	SO_3	6.0×10^{-3}
HCl	1.5×10^{-2}		

		Relative Saturations		
CaO (g moles/liter)	pH	$Ca(OH)_2$	$CaSO_3$	$Mg(OH_2)$
1.6×10^{-2}	5.7	1.1×10^{-12}	9.2	1.4×10^{-7}
1.7×10^{-2}	6.0	5.1×10^{-12}	14.8	6.2×10^{-7}
1.8×10^{-2}	6.3	2.0×10^{-11}	20.4	2.4×10^{-6}
1.9×10^{-2}	6.7	9.9×10^{-11}	26.0	1.2×10^{-5}
2.0×10^{-2}	7.3	1.8×10^{-9}	31.6	2.2×10^{-4}
2.1×10^{-2}	10.0	3.9×10^{-4}	34.2	41.7
2.2×10^{-2}	10.4	3.3×10^{-3}	34.9	279
2.3×10^{-2}	10.7	9.8×10^{-3}	35.5	690
2.4×10^{-2}	10.8	2.1×10^{-2}	35.9	1,230
2.5×10^{-2}	10.9	3.8×10^{-2}	36.3	1,870
2.6×10^{-2}	11.0	6.1×10^{-2}	36.6	2,580

TABLE 4-8

WORKING VOLUME OF THE LG & E SCRUBBING LOOP

Vessel	Volume (liters)
Marble beds (4)	28,700
Scrubber tanks (2)	204,000
Draft tube	12,000
Reaction tank	775,000
Surge tank	94,500
TOTAL*	1,114,200

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* Total excluding draft tube = 1,002,200 liters

If both scrubbers are operating at full capacity, roughly 400 g moles/min of SO_2 are removed from the flue gas. If an oxidation rate of 5% is assumed, 380 g moles/min must be precipitated as calcium sulfite. Therefore, up to 160 g moles/min of calcium sulfite could have been precipitated in the draft tube. If no nucleation occurs, this would require a driving force in the draft tube of over 100 times the driving force in the rest of the system. The maximum relative saturation calculated for the draft tube is around forty. Therefore, it is probable that significant amounts of nucleation take place in the draft tube.

From this analysis it is observed that the function of the draft tube may be more than one of mixing. A significant amount of nucleation probably occurs which increases the number of sites available for further precipitation. Almost half of the total precipitation for the system could occur in the draft tube.

4.6 Conclusions and Recommendations

The conclusions which can be drawn from this review of previous data include:

1) The oxidation of sulfite to sulfate is between five and ten percent.

2) The Paddy's Run No. 6 scrubbing unit operates in a subsaturated mode with respect to calcium sulfate. This is a result of both the low system sulfite oxidation level and the coprecipitation of calcium sulfate with calcium sulfite.

3) The consistency of the majority of the liquid phase analytical data gathered was reasonable as judged by electroneutrality balances using the Radian chemical equilibria program. However, charge imbalances in excess of 10% were calculated for some of the data. It was not possible to determine the cause of these imbalances without independent testing.

4) Several inconsistencies were noted when the data were used in precipitation rate calculatins. Analytical and/or sampling procedural errors were probably responsible for these inconsistencies.

5) Significant amounts of nucleation and precipitation probably occur in the draft tube.

There are several recommendations which might improve the results derived from future sampling and analysis. Short sample lines for scrubber effluent liquor samples and adequate flushing are required for good results. All points around the scrubber should be sampled simultaneously if possible to eliminate possible inconsistencies resulting from short scrubber residence times. Carbonate analyses should be performed on future samples.

In addition to sampling procedure improvement, additional sample points could be used to help define and check system performance. The downcomer lines could be sampled both immediately after exiting the scrubber and before entering the hold tank. This would provide an independent consistency test. Also, sampling around the draft tube would be beneficial for verification of draft tube chemistry.

5.0 COMPUTER SIMULATION OF THE LG&E LIME SCRUBBING SYSTEM

The Paddy's Run No. 6 lime scrubbing system was modeled using the Radian Process Simulation Model. It should be noted that pure phase kinetics were used in the simulation since the coprecipitation kinetics have not been defined. While this limits the scope of the study somewhat, some useful results were derived from the simulations. The major objectives of this simulation are listed below:

1) To compare simulated system parameters (primarily $CaSO_3 \cdot \frac{1}{2}H_2O$ relative saturations) with the actual data gathered at LG&E to further identify and analyze possible inconsistencies in the data.

2) To better understand the significant process chemistry steps in the important process vessels of the system. This involved varying the lime addition rate and calcium sulfite dissolution rate in an effort to more adequately model the actual LG&E system. The effect of changing these parameters on the performance of scrubber, draft tube, and hold tank was examined. The various streams were checked for possible scaling conditions.

This section provides a brief description of the Radian Process Model which was used. A more detailed subroutine documentation is furnished in Appendix II. A discussion of the required inputs and necessary assumptions is also supplied.

5.1 Process Simulation System Description

The Radian process simulation system is a collection of computer subroutines which are designed to model the performance characteristics of aqueous inorganic chemical processes. Since this package was originally developed to assist in the evaluation of aqueous SO_2 sorption processes, the advantages of its use for this program are obvious.

The calculations which are made in a scrubber simulation can generally be grouped into two categories: (1) material and energy balances around the scrubber system and (2) the prediction of vapor-liquid-solid equilibria. The simulation package itself is composed of an executive system and a set of equipment subroutines.

The function of the executive system is to interconnect the various process units in the appropriate fashion and to control the sequencing of computer operations. The executive system also interfaces the equipment subroutines with the computer input/output devices. Further explanation of the executive system is given in Appendix II.

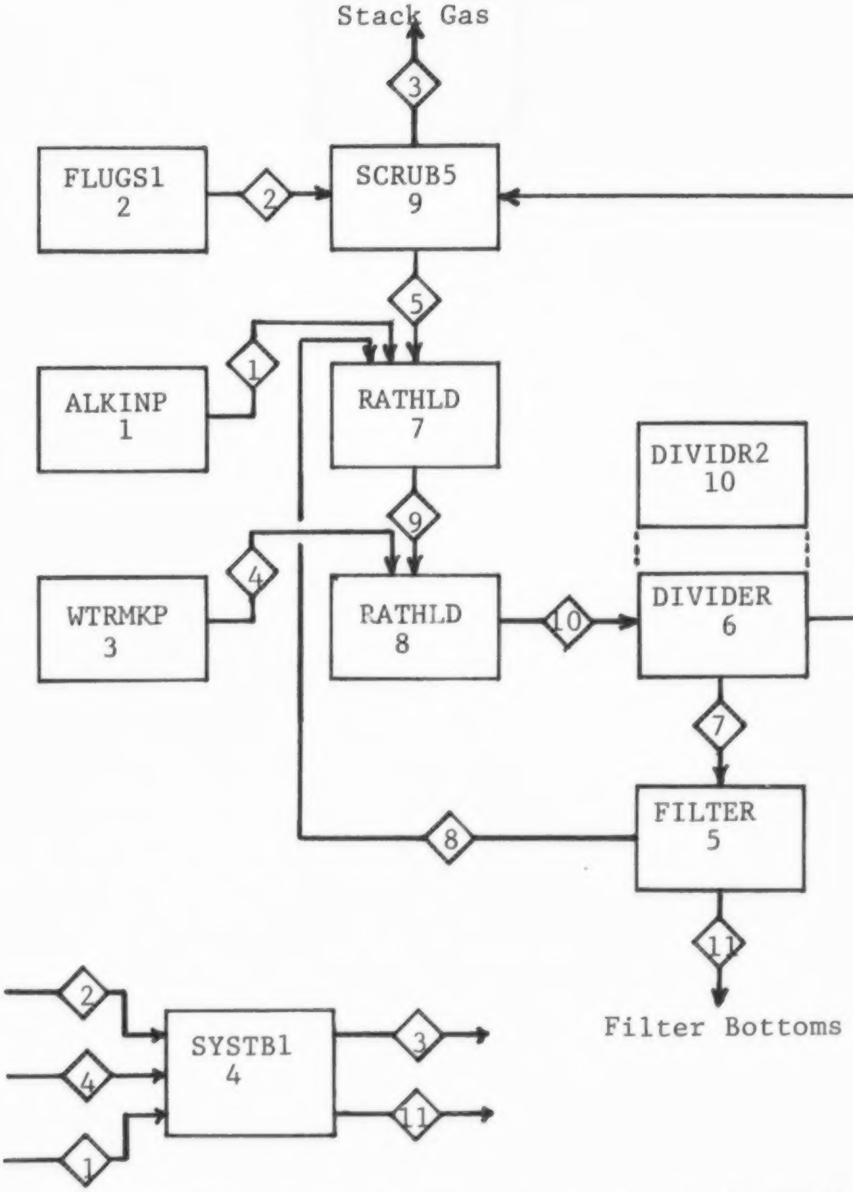
The functions of the equipment subroutines are to convert input data to the forms which are used within the programs and to perform unit operation calculations. Where necessary, these subroutines make use of the Radian chemical equilibrium program.

The overall computational scheme used in the model to simulate the performance of the LG&E desulfurization system is shown in Figure 5-1. The order of process calculations is also listed. There are three input equipment boxes. ALKINP (Equipment Box No. 1) specified the flow of alkali material into the system. FLUGS1 (Equipment Box 2) provided the input data for the flue gas stream entering the scrubber. WTRMKP (Equipment Box 3) specified the composition of the water make-up stream.

Once the inputs were specified and an initial estimate for the amount of CO_2 sorbed in the scrubber was made, the flow rate of the water make-up stream was calculated by material balance in SYSTB1 (Equipment Box 4). SYSTB1 also calculated the equilibrium composition and flow rate of the filter bottoms stream (the percent solids is specified) and the amount of water vapor in the stack gas.

After the filter bottoms stream had been specified by SYSTB1, subroutine FILTER (Equipment Box 5), a solid-liquid separator, completely specified the flow rates and compositions of the filter inlet and overflow streams. (FILTER is actually a model of both the clarifier and filter which exist at Paddy's Run.) These calculations were based on knowledge of solids loadings in all of the streams and the assumption that the filter inlet and exit streams were all in solid-liquid equilibrium. DIVIDER and DIVDR2 (Equipment Boxes 6 and 10) are stream splitters. DIVDER was used in conjunction with FILTER as an entry into the scrubbing loop. DIVDR2 was used thereafter.

Next, the scrubber was simulated as a backmixed scrubber by subroutine SCRUB5 (Equipment Box 9). The input streams to the scrubber were the scrubber feed stream (specified by DIVDER or DIVDR2) and the flue gas stream (specified by input


Stack Gas
3
FLUGS1
2
2
SCRUB5
9
5
1
ALKINP
1
RATHLD
7
6
DIVIDR2
10
9
4
WTRMKP
3
RATHLD
8
10
DIVIDER
6
7
FILTER
5
8
11
Filter Bottoms
2
SYSTB1
4
3
4
11
1

Order of Process Calculations: 1, 2, 3 (4, 5, 6 (9, 7, 8, 10))

FIGURE 5-1. LG & E PROCESS SIMULATION FLOWSHEET

routine FLUGS1). The composition of the stack gas was calculated by specifying fractional removals of SO_2 and CO_2 and by assuming adiabatic saturation. No solids precipitation was allowed in the scrubber. A specified fraction of the solids in the scrubber feed stream was allowed to dissolve in the scrubber. After the specified solid, liquid, and gas mass transfer reactions were accounted for, the Radian equilibrium program was used to calculate the composition of the scrubber bottoms stream.

The scrubber bottoms liquor, lime solids, and the filter overflow stream were the input streams to the first of two subroutines named RATHLD (Equipment Boxes 7 and 8). The first of these simulated the draft tube. The draft tube effluent was an input to the hold tank which was also modeled as subroutine RATHLD. RATHLD simulates a non-equilibrium hold tank. Calcium sulfate and sulfite supersaturations were allowed and precipitation rates of both of these species were calculated using established correlations for $CaSO_3/CsSO_4$ precipitation rates as functions of the relative saturations of precipitating ionic species. The rate constants used in these calculations were derived from Radian experimental data. Since the kinetics of $CaSO_4$ coprecipitation with $CaSO_3$ have not been defined, these rates were based on precipitating calcium sulfate only if its relative saturation exceeded one. Calcium carbonate was assumed to be at equilibrium since only a small amount is formed in the LG&E system. RATHLD was used twice so that the draft tube could be modeled separately from the hold tank and the surge tank.

The precipitating species solids compositions in hold tank effluent stream were compared with the values obtained for this stream during the preceding iteration. If the compositions for consecutive calculations were consistent, the solids concentrations and, therefore, the precipitation rates were converged.

If not, the "loop" (Equipment Boxes 10, 9, 7, 8) was repeated until the concentrations agree to within 10% for successive calculations. This convergence loop is required since the original estimate of the scrubber feed slurry composition was assumed to be in solid-liquid equilibrium with the filter effluent stream. The actual scrubber feed stream was not in equilibrium with the filter bottoms, and adjustments were required.

Once the precipitation rates were converged, the fraction of CO_2 absorbed from the flue gas was adjusted toward a value which was consistent with gas-liquid CO_2 equilibrium between the stack gas and the scrubber bottoms stream. Then the overall material balances were repeated and the precipitation rates were again converged. When the amount of CO_2 sorbed and the precipitation rates both met the criteria for convergence, the simulation case was ended. (If the CO_2 sorption fraction was within 10% of the equilibrium value, the CO_2 loop was considered to be converged.) An external convergence was required to adjust the amount of lime needed to approximate the pH of 8 in the filter bottoms at LG&E. (Note: The 8 pH is the result of a piping change since December, 1973. No actual analytical data for the filter bottoms is available for comparison with simulation results.)

5.2 Input Information and Key Assumptions

The specific input values used are listed in Appendix III with a sample simulation case. The composition and flow rate of the flue gas were derived from information gathered from Louisville Gas and Electric. The SO_2 level was an actual operating measurement. An SO_2 removal of 85% was assumed in the scrubber. The other numbers were derived from a typical coal analysis and air leakage figures. (The hydrogen chloride composition was calculated from the chloride content of the scrubbing liquor.) A complete analysis of the make-up water was provided by Combustion Engineering. Makeup lime was assumed to be almost pure calcium hydroxide; however, to obtain the magnesium ion concentration reported in the actual scrubbing liquor, the lime was also assumed to contain .06% wt magnesium hydroxide.

There are several assumptions which were required to perform the process simulations in this section. These are presented below:

1) The scrubber exit stack gas was assumed to be saturated with respect to water.

2) 85% SO_2 removal efficiency was assumed.

3) Since no solubility data exist for the solids entering the system with the flue gas (fly ash), no alkalinity was assumed to be derived from this source. (All particulates were assumed to be inert.)

4) Equilibrium exists between CO_2 in the stack gas and the scrubber exit liquor.

5) The scrubber exit liquor and gas temperatures were assumed to be equal to the adiabatic saturation temperature.

6) The scrubber was modeled without allowing solids precipitation to occur. However, by specifying an assumed fraction to be dissolved, dissolution of $Mg(OH)_2$, $Ca(OH)_2$, and $CaSO_3 \cdot \frac{1}{2}H_2O$ solids entering the scrubber was allowed.

7) The two scrubbers, each having two marble beds, were modeled as a single backmixed unit. This assumption is justified by the fact that the pH's of the liquid streams leaving both the upper and the lower bed in each scrubber are approximately equal. Since the gas sees essentially the same liquor on both beds, the two bed scrubber used performs as if it were a single bed scrubber having a packed bed twice as thick as one of the existing beds.

8) Calcium carbonate was an equilibrium species in the rate hold tanks. Since there is little carbonate formed, the error derived from this assumption is also small.

9) The filter subroutine assumes that the input and output streams are in chemical equilibrium. (This subroutine is actually a model of the clarifier and vacuum filter system which exist at LG&E.)

10) All oxidation was assumed to occur in the scrubber.

11) Conventional, pure phase kinetics were used. Calcium sulfate was allowed to precipitate only if its equilibrium solubility was exceeded. Coprecipitation was not considered since the kinetics of this phenomenon have not been determined.

5.3 Discussion of Results

The results of the Radian simulation of the LG&E lime scrubbing system are presented in this subsection. Several significant aspects of these results are discussed in separate subsections below.

5.3.1 Lime Addition

The quantity of lime which was specified as being added to the hold tank was varied in an attempt to duplicate the hold tank effluent pH of 8 which has been reported at Paddy's Run. The simulation results are listed in capsule form in Table 5-1. Detailed simulation results for one of the runs is shown in Appendix III. It was found that the simulated hold tank effluent pH was insentive to lime addition rate in the range of 69.8 to 70.2 pounds per minute with the pH remaining between 6.6 and 6.8 for these cases (100.7 to 101.3% stoichiometry). However, an increase to 70.4 pounds of lime added per minute (101.6% stoichiometry) increased the hold tank effluent pH to 9.2.

These results indicate that the hold tank effluent pH is a very sensitive function of lime addition rate in the pH range of 7-9 (a very desirable feature from a process control point of view). Calcium sulfite and sulfate supersaturations (and hence their calculated precipitation rates) however, did not vary significantly with this increase in lime addition rate (and pH). For this reason, the 70.2 lb/min lime addition rate was chosen for further simulation studies.

5.3.2 Alkalinity Required in the Scrubber

In order to scrub a specified amount of acidic species from the gas entering the scrubber, sufficient basic species must be available in the scrubber feed. The liquid flow rate required to scrub a specified amount of SO_2 is dependent upon the concentrations of available basic materials in the scrubber feed.

TABLE 5-1

SIMULATION RESULTS SUMMARY

			Relative Saturation			Precipitation Rates (g moles/sec)	
	pH	P_{SO_2} * (atm)	Calcium Sulfate	Calcium Sulfite	Calcium Hydroxide	Calcium Sulfate	Calcium Sulfite
Case 1							
Scrubber Effluent	2.6	4.3×10^{-3}	0.91	5.2×10^{-3}	9.7×10^{-19}	-	-
Draft Tube Effluent	7.0	2.5×10^{-8}	1.22	21.2	5.5×10^{-10}	3.7×10^{-3}	0.9
Hold Tank Effluent	6.8	6.1×10^{-9}	1.12	1.8	1.9×10^{-10}	0.3	5.8
Case 2							
Scrubber Effluent	2.6	4.2×10^{-3}	1.03	6.2×10^{-3}	9.7×10^{-19}	-	-
Draft Tube Effluent	7.0	2.6×10^{-8}	1.21	21.2	5.4×10^{-10}	3.7×10^{-3}	0.9
Hold Tank Effluent	6.7	6.2×10^{-9}	1.12	1.8	1.9×10^{-10}	0.3	5.8
Case 3							
Scrubber Effluent	2.8	2.6×10^{-3}	1.14	1.2×10^{-2}	3.0×10^{-18}	-	-
Draft Tube Effluent	9.1	1.3×10^{-12}	1.22	23.6	1.2×10^{-5}	3.6×10^{-3}	1.0
Hold Tank Effluent	9.1	1.2×10^{-13}	1.13	1.8	9.7×10^{-5}	0.3	5.7
Case 4							
Scrubber Effluent	5.6	9.9×10^{-6}	1.40	20.0	1.3×10^{-12}	-	-
Draft Tube Effluent	7.0	4.6×10^{-8}	1.27	56.6	8.1×10^{-10}	4.6×10^{-3}	2.3
Hold Tank Effluent	6.6	1.8×10^{-8}	1.11	3.0	1.2×10^{-10}	0.2	15.2
Case 5							
Scrubber Effluent	3.8	5.7×10^{-4}	1.36	0.3	3.4×10^{-16}	-	-
Draft Tube Effluent	7.1	2.8×10^{-8}	1.26	39.9	9.3×10^{-10}	4.3×10^{-3}	1.6
Hold Tank Effluent	6.7	9.0×10^{-9}	1.13	2.4	1.7×10^{-10}	0.3	10.5

	Lime Rate (lb/min)	Filter Bottoms pH	Sulfite Dissolution in Scrubber (%)	L/G (gal/acf)
Case 1	69.8	6.8	0.0	34.0
Case 2	70.2	6.8	0.0	34.0
Case 3	70.4	9.1	0.0	34.0
Case 4	70.2	6.6	2.0	34.0
Case 5	70.2	6.7	1.0	34.0

Normally, a liquid is considered basic (or alkaline) when the pH is high (denoting a high concentration of OH^-). For aqueous scrubbing of SO_2 (the acid anhydride of sulfurous acid), a number of other liquid species should be considered basic. This means that these species are basic with respect to the first ionization step of sulfurous acid, i.e.,

$$H_2SO_3(\ell) \rightarrow HSO_3^- + H^+ \qquad (5\text{-}1)$$

Of the chemical species present in a lime scrubbing system, the following species are basic with respect to SO_2 sorption:

OH^-
$SO^{3=}$
HCO_3^-
$CO_3^=$

$CaOH^+$
$CaSO_3(\ell)$
$CaCO_3(\ell)$
$CaHCO_3^+$

$MgOH^+$
$MgSO_3(\ell)$
$MgHCO_3^+$
$MgCO_3(\ell)$

$NaOH(\ell)$
$NaCO_3^-$
$NaHCO_3(\ell)$

In summing the basic species which are available in the scrubber feed, most of the species in solution are given a weighting factor of one. The carbonate species $[CO_3^=$, $CaCO_3(\ell)$, $MgCO_3(\ell)$, $NaCO_3^-]$ have been given a weighting factor of two since they can react with two hydrogen ions.

There are essentially two acidic species present in scrubber liquor. A given fraction of the SO_2 absorbed is specified as oxidizing to SO_3 on absorption. The SO_3 is twice as acidic as the SO_2 with respect to the first ionization of sulfurous acid. This is because the sulfuric acid disassociates far more extensively than does sulfurous acid.

$$SO_3 + H_2O \rightarrow H_2SO_4(\ell) \rightarrow HSO_4^- + H^+ \rightarrow SO_4^= + 2H^+ \qquad (5\text{-}2)$$

Thus, SO_3 has been given a weighting factor of two.

During the initial LG&E simulation runs, a lack of sufficient alkalinity in the scrubber feed was noted. Even though enough lime had been added to exceed the hold tank effluent pH which was sought, sufficient basicity was not present to achieve the specified level of sulfur dioxide removal (85%). Indications of low scrubber alkalinity were noted in the pH and SO_2 equilibrium partial pressure values of the scrubber bottoms stream. The calculated pH was from 2.5 to 3 instead of being approximately 5. The computed equilibrium partial pressure of SO_2 over the scrubber bottoms was higher than the specified partial pressure in the stack gas (4.3×10^{-4} atm). This was obviously not realistic.

Since not enough liquid phase alkalinity was available in the scrubber feed liquor, some basicity had to be provided by solids dissolution in the scrubber. As can be determined from the calcium hydroxide relative saturations listed in Table 5-1, there was not solid phase calcium hydroxide in the hold tank effluent which was capable of providing this alkalinity since, in these simulations, all of the calcium hydroxide was assumed to be dissolved in the draft tube. Since solid $Ca(OH)_2$ could therefore not be the species providing the necessary scrubber solid phase alkalinity, some $CaSO_3$ dissolution was assumed.

It should be noted that the scrubber bottoms stream was extremely subsaturated with respect to calcium sulfite in the first three simulation cases. The analytical data gathered at Paddy's Run shows that this stream was actually slightly supersaturated. Furthermore, rate calculations using Paddy's Run data indicated that a significant quantity of calcium sulfite dissolved in the scrubber. (This is obviously a contradiction in the data since calcium sulfite dissolution cannot occur in a liquor which is supersaturated with respect to calcium sulfite.)

An acid-base balance using simulation results indicated that between one and two percent sulfite dissolution was required for the additional alkalinity needed to obtain a scrubber effluent pH of approximately 5.

Dissolution of two percent of the calcium sulfite solids in the scrubber furnished sufficient alkalinity for the specified SO_2 removal (Case 4). The scrubber bottoms pH was 5.6 and the equilibrium SO_2 partial pressure for this stream was 9.9×10^{-6} atm. However, further investigation of this case indicates that two percent $CaSO_3 \cdot \frac{1}{2}H_2O$ dissolution is not realistic. The calcium sulfite relative saturation in the scrubber bottoms stream was twenty (refer to Table 5-1). As discussed previously, there is no driving force for dissolution when the relative saturation exceeds one.

Specifying a dissolution rate of one percent of the calcium sulfite hemihydrate solids entering the scrubber resulted in a scrubber bottoms pH of 3.8. The equilibrium SO_2 partial pressure for this stream was 5.7×10^{-4} atm. This slightly exceeds the specified SO_2 partial pressure in the stack gas (4.3×10^{-4} atm). The relative saturation for the scrubber bottoms stream was 0.29 for this case. It is seen by comparison of the last two cases that the amount of dissolution required to adequately model the LG&E system is between one and two percent. However, until more is known about the kinetics of sulfite dissolution in the scrubber, an expression for the extent of dissolution cannot be established. The benefits which would be derived from further trial and error determination of the extent of sulfite dissolution do not justify the increased cost of the computer time required.

It should be noted that all of the simulated streams were calculated to be supersaturated with respect to calcium sulfate dihydrate. These $CaSO_4 \cdot 2H_2O$ supersaturations resulted from utilization of pure phase rate kinetics rather than coprecipitation kinetics since coprecipitation kinetics could not be modeled. When the pure phase kinetic model is employed, calcium sulfate solids can be precipitated only if calcium sulfate supersaturations occur. Although calcium sulfate supersaturations are not in evidence at Paddy's Run, this shortcoming should not significantly affect other simulation results which were discussed previously.

In summary, one simulation case (Case 5) approximates reasonably well the operation of the lime scrubbing system at Paddy's Run No. 6 (considering that coprecipitation cannot be modeled at this time). This case indicates that some calcium sulfite dissolution in the scrubber (1 to 2%) is required to provide sufficient alkalinity for 85% SO_2 removal.

5.4 Conclusions

The conclusions reached during the simulation of the LG&E system are listed below.

1) To completely and adequately model the LG&E scrubbing system, more must be known about fundamental nature of the coprecipitation phenomenon. An understanding of the factors which lead to a low sulfite oxidation rate is also of critical importance since it is the combination of these factors which make it possible for the LG&E system to operate subsaturated with respect to calcium sulfate.

2) In the 7-9 pH range, the simulated hold tank pH is extremely sensitive to the lime addition rate. Other important rates and saturation levels do not fluctuate significantly as the lime addition rate is changed.

3) The importance of calcium sulfite dissolution in the scrubber is emphasized. The sulfite must dissolve to furnish the alkalinity required for good scrubber performance. However, dissolution kinetic data is needed to adequately predict the amount of dissolution which occurs.

6.0 APPENDICES

6.1 APPENDIX I - SAMPLE RATE CALCULATIONS

TABLE 6-1

SAMPLE RATE CALCULATIONS FOR SCRUBBER

Date	Vessel	Reaction	Method of Calculation		Stream Name	Stream Flow Rate	Total Species in Liquid (m moles/liter)	Total Species in Gas (ppm)	Total Species Flow Rate (g moles/min)
11/23/73	'B' Scrubber	$CaSO_3 \cdot \frac{1}{2}H_2O$ precipitation	Overall SO_2 Balance-oxidation	Incoming	Flue gas inlet	129,100 g moles/min		1,750	225.9
					Scrubber feed	22,680 1/min	.79		17.9
				Outgoing	Flue gas outlet	135,600 g moles/min		230	31.2
					Upper bed downcomer	7,560 1/min	9.0		68.0
					Lower bed downcomer Scrubber bottoms	15,120 1/min	12.3		186.0
		$CaSO_4 \cdot 2H_2O$ precipitation	Liq. SO_3 Balance + oxidation	Incoming	Scrubber feed	22,680 1/min	6.3		142.9
				Outgoing	Upper bed downcomer	7,560 1/min	5.2		39.3
					Lower bed downcomer Scrubber bottoms	15,120 1/min	5.3		80.1
		$Ca(OH)_2$ Dissolution	Liq. Ca Balance + Σ ppt	Incoming	Scrubber feed	22,680 1/min	10.7		242.7
				Outgoing	Upper bed downcomer	7,560 1/min	13.8		104.6
					Lower bed downcomer Scrubber bottoms	15,120 1/min	15.1		228.3

$$CaSO_3 \cdot \tfrac{1}{2}H_2O \text{ ppt rate} = \Sigma\, SO_2(aq + g)_{in} - \Sigma\, SO_2(aq + g)_{out} - \text{oxidation rate}$$

$$= 243.8 - 285.2 - (.06)(194.7) = -53.1 \text{ g moles/min}$$

$$CaSO_4 \cdot 2H_2O \text{ ppt rate} = \Sigma\, SO_3(aq)_{in} - \Sigma\, SO_3(aq)_{out} + \text{oxidation rate}$$

$$= 142.9 - 119.4 = 11.7 = 35.2 \text{ g moles/min}$$

$$Ca(OH)_2 \text{ dissolution rate} = \Sigma\, Ca(aq)_{out} - \Sigma\, Ca(aq)_{in} + \Sigma \text{ ppt}$$

$$= 332.9 - 242.7 - 17.9 = 72.3 \text{ g moles/min}$$

TABLE 6-1 - SAMPLE RATE CALCULATIONS FOR SCRUBBER (Cont.)

Date	Vessel	Reaction	Method of Calculation		Stream Name	Stream Flow Rate	Total Species in Liquid (m moles/liter)	Total Species in Gas (ppm)	Total Species Flow Rate (g moles/min)
11/23/73	Hold Tank	$CaSO_3\cdot\frac{1}{2}H_2O$ precipitation	Liquid SO_2 Balance	Incoming	Upper 'A' downcomer	-			
					Upper 'B' downcomer	7,560 l/min	9.0		68.0
					Lower 'A' downcomer	-			
					Lower 'B' downcomer	11,340 l/min	12.3		139.5
					'A' scrubber bottoms	-			
					'B' scrubber bottoms	3,780 l/min	12.3		46.5
				Outgoing	Hold tank effluent	22,680 l/min	0.79		17.9
				$CaSO_3\cdot\frac{1}{2}H_2O$ ppt rate $= \Sigma\ SO_2(aq)_{in} - \Sigma\ SO_2(aq)_{out}$ $= 254.0 - 17.9 = 236.1$ g moles/min					
		$CaSO_4\cdot 2H_2O$ precipitation	Liquid SO_4 Balance	Incoming	Upper 'A' downcomer	-			
					Upper 'B' downcomer	7,560 l/min	5.2		39.3
					Lower 'A' downcomer	-			
					Lower 'B' downcomer	11,340 l/min	5.3		60.1
					'A' scrubber bottoms	-			
					'B' scrubber bottoms	3,780 l/min	5.3		20.0
				Outgoing	Hold tank effluent	22,680 l/min	6.3		142.9
				$CaSO_4\cdot 2H_2O$ ppt rate $= \Sigma\ SO_4(aq)_{in} - \Sigma\ SO_4(aq)_{out}$ $= 119.4 - 142.9 = 23.5$ g moles/min					
		$Ca(OH)_2$ dissolution	Liquid Ca Balance + Σ ppt	Incoming	Upper 'A' downcomer	-			
					Upper 'B' downcomer	7,560 l/min	13.8		104.6
					Lower 'A' downcomer	-			
					Lower 'B' downcomer	11,340 l/min	15.1		171.2
					'A' scrubber bottoms	-			
					'B' scrubber bottoms	3,780 l/min	15.1		57.1
				Outgoing	Hold tank effluent	22,680 l/min	10.7		242.7
				$Ca(OH)_2$ dissolution rate $= \Sigma\ Ca(aq)_{out} - \Sigma\ Ca(aq)_{in} + \Sigma$ ppt $= 242.7 - 332.9 + 212.6 = 122.4$ g moles/min					

6.2 APPENDIX II - SUBROUTINE DOCUMENTATION

EXECUTIVE SYSTEM

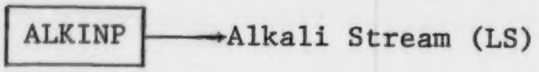
The function of the executive system is to interconnect the various units in the appropriate fashion and control the sequencing of the computer operations.

The process units are interconnected by means of a process matrix during an initialization phase of computer operations. In this phase, model input data are read into the machine, the process matrix is used to define the processing scheme, and each equipment box is initialized. Each processing unit is labeled by a number called an equipment number and the subroutine designation. The various process streams are labeled by a stream number.

The process matrix used to define each processing scheme is given on the first page of computer printout for each simulation case. Each process unit is listed in the process matrix. Input and output stream vectors are assigned to each unit so that the interconnections specified in the process matrix correspond to the interconnections of the process flow diagram.

The executive system also must be given the order in which the process calculations are to be made. This order is indicated immediately under the process matrix. To execute the process calculations, the executive system takes each subsequent equipment number from the order of calculation and determines the subroutine name and the input and output streams from the process matrix.

ALKINP: ALKALI INPUT ROUTINE



Function: This routine simulates a limestone stream input to the process on the basis of specified flow rate and composition and if desired, temperature.

Input Information: The flow rate of the alkali stream ($\hat{W}_{LS,s}$, lb/min) must be specified, along with the weight fraction composition ($XW_{LS,js}$) of the following species: (1) CaO, (2) MgO, (3) $CaCO_3$, (4) $MgCO_3$, (5) $CaSO_3$, (6) $MgSO_3$, (7) $CaSO_4$ (8) $MgSO_4$, (9) inert, (10) $Ca(OH)_2$, (11) $Mg(OH)_2$.

Output Information: The alkali stream will be completely specified.

Description: The first step in the calculations for this routine is to convert the weight flow rate of the alkali input stream from input units ($\hat{W}_{LS,s}$, lb/min) to program units ($W_{LS,s}$, g/sec).

$$W_{LS,s} = 453.5924\ \hat{W}_{LS,s}/60 \qquad (1)$$

The molar flow rates of the various components are now determined by Equation 2.

$$N_{js} = W_{LS,s} xw_{LS,js}/MW_{js} \qquad (2)$$

One additional parameter is required to describe the input stream: the mole fraction of magnesium carbonate relative to the total carbonate containing species ($MgCO_3$ plus $CaCO_3$) in the alkali stream (xl_{Mg}), as defined in Equation 3.

$$xl_{Mg} = \frac{\dfrac{W_{LS,s}xw_{LS,MgCO_3}}{MW_{MgCO_3}}}{\dfrac{W_{LS,s}xw_{LS,MgCO_3}}{MW_{MgCO_3}} + \dfrac{W_{LS,s}xw_{LS,CaCO_3}}{MW_{CaCO_3}}} \qquad (3)$$

Equation 3 may be simplified as in Equation 4.

$$xl_{Mg} = \frac{xw_{LS,MgCO_3}MW_{CaCO_3}}{xw_{LS,MgCO_3}MW_{CaCO_3} + xw_{LS,CaCO_3}MW_{MgCO_3}} \qquad (4)$$

If this parameter is greater than 0.5, the simulation is terminated. The magnesium carbonate never exceeds the calcium carbonate content in limestone, and, if the computed ratio is larger than 0.5, an input error is indicated.

The last step in the calculations for this routine is to convert the limestone stream temperature from input units ($\hat{T}_{LS}$, °F) to program units (T_{LS}, °K) if the temperature is specified.

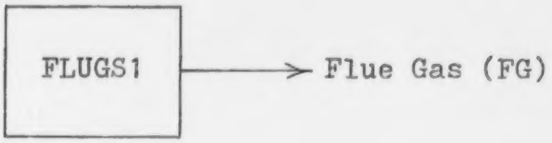
$$T_{LS} = 273.16 + (\hat{T}_{LS} - 32)/1.8 \qquad (5)$$

The temperature is set at 273.16°K if it is not specified.

Nomenclature:

LS	designation for the limestone stream
MW	molecular weight
n	molar flow rate of a species or component
T	temperature
W	weight flow rate of a stream of phase
xl_{Mg}	fraction of magnesium in limestone
xw	weight fraction of a species or component
^	indicates a quantity not in program units

FLUGS1: FLUE GAS INPUT ROUTINE



Function: This routine simulates a flue gas stream with entrained fly ash on the basis of specified temperature, pressure, flow rate, and composition of the flue gas, and specified flow rate and composition of the fly ash.

Input Information: The temperature, pressure, and flow rate of the flue gas ($\hat{T}_{FG}$, °F, $\hat{P}_{FG}$, psia, and $\hat{V}_{FG}$, acfm) must be specified, along with the mole fraction composition ($y_{FG,jg}$, for jg = all gas species). The flow rate of the fly ash ($\hat{W}_{FA,s}$, lb/min) must be specified, along with the weight fraction composition ($xw_{FA,js}$, js = all fly ash species). The species which are considered in the fly ash are CaO, $CaSO_4$, MgO, Na_2O, NaCl, and inerts.

Output Information: The flue gas stream is completely specified.

Assumption: All of the species in the flue gas behave ideally.

Description: The first step in the calculations for this routine is to convert the temperature and pressure from input units ($\hat{T}_{FG}$, ^{o}F and $\hat{P}_{FG}$, psia) to program units (T_{FG}, ^{o}K and P_{FG}, atm).

$$T_{FG} = 273.16 + (\hat{T}_{FG} - 32)/1.8 \qquad (1)$$

$$P_{FG} = \hat{P}_{FG} / 14.696 \qquad (2)$$

The next step in the calculations is to find the total molar rate of the flue gas ($N_{FG,g}$, gmole/sec). This is done by applying the ideal gas law, which is given in common nomenclature in Equation 3.

$$P V = n R T \qquad (3)$$

Equation 4 is a rearrangement of Equation 3 in which the nomenclature has been changed and the proper conversion factors have been added.

$$N_{FG,g} = \frac{28{,}316.85 \ \hat{V}_{FG} \ P_{FG}}{82.0575 \cdot 60 \ T_{FG}} \qquad (4)$$

The molar flow rates of the individual gas species in the flue gas ($n_{FG,jg}$, jg = all gas species) may now be found by multiplying the total molar flow rate ($N_{FG,g}$, gmole/sec) by the mole fractions ($y_{FG,jg}$, jg = all gas species).

$$n_{FG,jg} = N_{FG,g} \; y_{FG,jg} \qquad (5)$$

for jg = all gas species.

The weight rate of the fly ash is converted from input units ($\hat{W}_{FA,s}$, lb/min) to program units ($W_{FA,s}$, g/sec) by using Equation 6.

$$W_{FA,s} = 453.5924 \; \hat{W}_{FA,s} / 60 \qquad (6)$$

The molar rates of the individual fly ash solids entrained in the flue gas ($n_{FG,js}$, js = all solid species) may now be calculated by multiplying the weight rate of the solids by the weight fraction of each solid ($xw_{FA,js}$) and dividing by its molecular weight (MW_{js}).

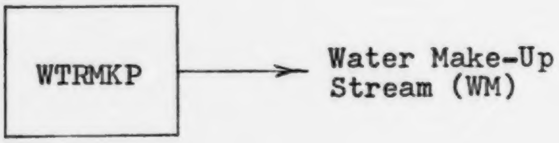
$$n_{FG,js} = W_{FA,s} \; xw_{FA,js} / MW_{js} \qquad (7)$$

for js = all solid species. This concludes the calculations for this routine.

Nomenclature:

FA	designation for the fly ash
FG	designation for the flue gas stream
MW	molecular weight
n	molar flow rate of a species or component
N	molar flow rate of a stream or phase
P	pressure
T	temperature
V	volumetric flow rate of a stream or phase
W	weight flow rate of a stream or phase
xw	weight fraction of a species or component
y	gas mole fraction
^	indicates a quantity not in program units

WTRMKP: WATER MAKE-UP ROUTINE



Function: This routine simulates a process water make-up stream on the basis of composition and if desired, temperature. The impurities which may be present in this stream (WM) are: $SO_3^=$, $CO_3^=$, $SO_4^=$, NO_3^-, Ca^{++}, Mg^{++}, Na^+, and Cl^-.

Input Information: The concentrations of the impurities must be specified ($c_{WI,jl}$, $jl = SO_3^=, CO_3^=, SO_4^=, NO_3^-, Ca^{++}, Mg^{++}, Na^+$, and Cl^-, mg/l). These impurity concentrations are taken to refer to the total species amount of the impurity, i.e., total sulfite as $SO_3^=$, total carbonate as $CO_3^=$, etc.

The temperature of the water make-up stream ($\hat{T}_{WM}$, ^oF) may be specified if desired.

Output Information: The temperature and the concentrations of the key ions in the make-up water stream are calculated in program units.

Description: The first step in the calculations for this routine is to convert the temperature of the water make-up stream from input units ($\hat{T}_{WM}$, ^oF) to program units (T_{WM}, ^oK).

$$T_{WM} = 273.16 + (\hat{T}_{WM} - 32.0) / 1.8 \tag{1}$$

If the temperature of the water make-up stream is undefined, that temperature is set to 323.16 ^{o}K (50 ^{o}C).

The next step in the calculations is to determine the molalities of the key ions ($m_{WI,jl}$, for jl = impurity ions, gmole/kg liq H_2O).

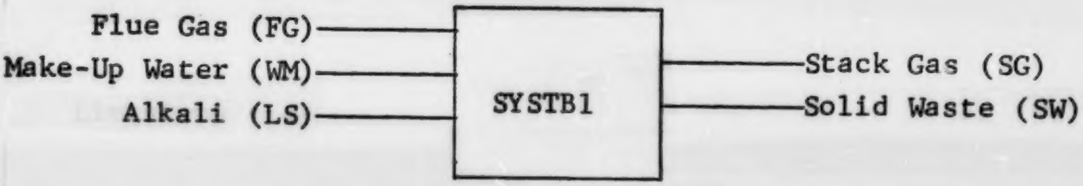
$$m_{WI,jl} = \frac{c_{WI,jl}}{\rho_{H_2O(l)}(T_{WI}) \ MW_{jl}} \tag{2}$$

for jl = all impurity ions, where $\rho_{H_2O(l)}(T_{WI})$ is the density of liquid water at the temperature of the water make-up stream (in g/l) and MW_{jl} is the molecular weight of one of these species. This completes the calculations for this routine.

Nomenclature:

c	concentration of a liquid species
m	molality of a liquid species
MW	molecular weight
T	temperature
WM	designation for the water make-up stream
ρ	density
^	indicates a quantity not in program units

SYSTB1: SYSTEM BALANCE ROUTINE



Function: This routine performs an overall heat and material balance around a scrubbing system. Based on complete information for the flue gas (FG) and alkali (LS) streams, the compositions of the make-up water (WM), and several parameters and assumptions, the flow rates of the make-up water stream will be calculated along with complete information for the stack gas (SG) and solid waste (SW).

Input Information: The flue gas and alkali streams must be completely specified. The compositions of the make-up water stream must be specified. The following parameters must be set.

(1) Fractions of SO_2, NO, and NO_2 absorbed from the flue gas (xa_{jg}, jg = SO_2, NO, and NO_2).

(2) Fraction of CO_2 absorbed from the flue gas (xa_{CO_2}). (If this parameter is not set it will be estimated as discussed below.)

(3) Fraction of the SO_2 absorbed which is oxidized to SO_3 in the scrubber (xo_s).

(4) Fractions of the solid CaO and MgO entering the system which are hydrated (x/su_{js}, js = CaO and MgO).

(5) Variables γ_1, γ_2, and γ_3 which are parameters used for pressure drop calculations.

(6) Variables β_1, β_2 and β_3 which are parameters for determining particulate removal.

(7) Solid waste pH criteria: median (pH_{MED}) and tolerance (pH_{TOL}). [If the pH median is not specified (i.e., equal to zero) no manipulation of the solid waste pH will be attempted.]

(8) Diameter of the scrubber (dia).

(9) Weight fraction of solids in the solid waste stream.

Output Information: The flow rate of the make-up water stream will be calculated. The stack gas and solid waste streams will be completely specified.

Assumptions:

(1) All of the SO_3 and HCl in the flue gas will be absorbed.

(2) The oxygen required to oxidize the NO and NO_2 absorbed to N_2O_5 and the SO_2 which is oxidized to SO_3 in the scrubber will be absorbed from the flue gas.

(3) All process streams except the inputs (flue gas, water make-up, and alkali) are at the adiabatic saturation temperature.

(4) The stack gas is saturated with water vapor.

(5) No vapor-liquid mass transfer occurs in the hold tank.

Description: The first step in the calculations for this routine is to determine if the fraction of CO_2 absorbed in the scrubber (xa_{CO_2}) has been given a value. Since this quantity may have both positive and negative values (for lime and limestone scrubbing systems, respectively) the test which is used is to see if the absolute value of xa_{CO_2} is equal to zero. If xa_{CO_2} is defined, operations proceed to the next step. If not, the value of xa_{CO_2} for a limestone scrubbing system is estimated.

In a lime scrubbing system CO_2 will be absorbed from the flue gas to form $CaCO_3$. The partial pressure of CO_2 in the flue gas will approach the CO_2 equilibrium partial pressure above the scrubber bottoms.

In a limestone scrubbing system, the scrubbing liquor will have an equilibrium partial pressure of CO_2 larger than the partial pressure of CO_2 in the flue gas so that CO_2 will be desorbed rather than absorbed. Radian has found through experience with calculations of this type that a good thumb rule for estimating the amount of CO_2 desorbed is to equate the CO_2 desorbtion rate of acidic gas species on an equal valence basis. For example, it is assumed that for each mole of SO_2 absorbed one mole of CO_2 will be desorbed, but for each mole of HCl absorbed one-half mole of CO_2 will be desorbed. This thumb rule is stated mathematically in Equation 1.

$$\begin{aligned} -n_{FG,CO_2} xa_{CO_2} &= n_{FG,SO_2} xa_{SO_2} + \tfrac{1}{2} n_{FG,NO} \cdot xo_{NO} \\ &\quad + \tfrac{1}{2} n_{FG,NO_2} xa_{NO_2} + n_{FG,SO_3} + \tfrac{1}{2} n_{FG,HCl} \end{aligned} \tag{1}$$

where the $n_{FG,jg}$ are the molar rates (gmole/sec) of the gas species in the flue gas and the xa_{jg} are the fractions of gas species which are absorbed. The minus sign accounts for the fact that CO_2 is desorbed rather than absorbed. Since all the other quantities are known, Equation 1 may be solved for xa_{CO_2}.

$$\begin{aligned} xa_{CO_2} = -\Big(& n_{FG,SO_2} xa_{SO_2} + \tfrac{1}{2} n_{FG,NO} xa_{NO} + \tfrac{1}{2} n_{FG,NO_2} xa_{NO_2} \\ & + n_{FG,SO_3} + \tfrac{1}{2} n_{FG,HCl} \Big) \end{aligned} \tag{2}$$

The next step in the calculations for this routine is to determine the pressure in the stack gas (P_{SG}, atm). First the pressure drop is calculated by Equation 3.

$$([L/G]^{\gamma_3})$$

$$\Delta\hat{P} = [\gamma_1 U_T^{\gamma_2}][(L/G)^{\gamma_3}] \tag{3}$$

where

$\Delta\hat{P}$ = pressure drop in psi
U_T = linear gas velocity through the scrubber (ft/sec)
L/G = liquid to gas ratio for the system (gpm/1000 acfm)
$\gamma_1, \gamma_2, \gamma_3$ = experimental parameters

The stack gas pressure is calculated by Equation 4.

$$P_{SG} = P_{FG} - \Delta\hat{P}/14.696 \tag{4}$$

where P_{FG} is the flue gas pressure and 14.696 is a conversion factor (psia/atm).

Next, the particulate removal is computed.

$$\text{xpart} = 1 - \exp[-\beta_1 (L/G)^{\beta_2} (U_T)^{\beta_3}] \tag{5}$$

where

xpart = fraction of particulates removed
$\beta_1, \beta_2, \beta_3$ = experimentally determined parameters

The next portion of this routine is concerned with the evaluation of the gas phase of the stack gas. At this point in the execution subroutine GABSO is called. Although this routine performs several calculations, the result of interest at this time is the overall rate at which gas species other than water are absorbed from the flue gas, $\Delta_a N_{FG,g}$ (gmole/sec). This overall absorbtion rate includes all the acid gas species (SO_2, CO_2, NO, NO_2, SO_3, and HCl) and the oxygen necessary to oxidize the NO and NO_2 to N_2O_5 and the SO_2 which oxidizes in the scrubber to SO_3. The oxidation reactions are shown in Equations 6 through 8 and $\Delta_a N_{FG,g}$ is defined in Equation 9.

$$2NO + \tfrac{3}{2} O_2 \rightarrow N_2O_5 \tag{6}$$

$$2NO_2 + \tfrac{1}{2}O_2 \rightarrow N_2O_5 \tag{7}$$

$$SO_2 + \tfrac{1}{2}O_2 \rightarrow SO_3 \tag{8}$$

$$\begin{aligned}\Delta_a N_{FG,g} = {} & n_{FG,SO_2} xa_{SO_2}(1.0 + \tfrac{1}{2}xo_S) + n_{FG,CO_2} \cdot xa_{CO_2} \\ & + n_{FG,NO} xa_{NO}(1.0 + \tfrac{3}{4}) + n_{FG,NO_2} \cdot xa_{NO_2}(1.0 + \tfrac{1}{4}) \\ & + n_{FG,SO_3} + n_{FG,HCl}\end{aligned} \tag{9}$$

Subroutine WAVAP in now called. Based on complete information for the flue gas stream, the overall absorbtion rate $\Delta_a N_{FG,g}$, and the stack gas pressure P_{SG} this routine performs a simultaneous heat and material balance to determine the adiabatic saturation temperature (T_{AS}, °K) and the rate of vaporization of water from the scrubber bottoms ($\Delta_v n_{SB,H_2O(\ell)}$, gmole/sec).

The mole fraction of water in the stack gas (y_{SG,H_2O}) is now calculated as the ratio of the vapor pressure of water at the adiabatic saturation temperature [$vp_{H_2O}(T_{AS})$, atm] to the total pressure of the stack gas (P_{SG}, atm) multiplied by the activity of liquid water in the scrubber bottoms [$a_{SB,H_2O(\ell)}$].

$$y_{SG,H_2O} = a_{SB,H_2O(\ell)} vp_{H_2O}\left(T_{AS}\right)/P_{SG} \tag{10}$$

If a value of the activity of water is available it is used. If not, the activity is assumed to be unity. The vapor pressure of water is defined empirically as in Equation 11.

$$vp_{H_2O}\left(T_{AS}\right) = 10\left(5.22684-1750.286/\left(T_{AS}-38.2\right)\right) \qquad (11)$$

The rate of vaporization of water from the scrubber bottoms is now calculated based on the mole fraction of water in the stack gas.

$$\Delta_v n_{SB,H_2O(\ell)} = \left(N_{FG,g} - \Delta_a N_{FG,g} - n_{FG,H_2O}\right) \cdot \frac{y_{SG,H_2O}}{1.0 - y_{SG,H_2O}} - n_{FG,H_2O} \qquad (12)$$

where $N_{FG,g}$ is the total gas phase flow rate of the flue gas (gmole/sec) and n_{FG,H_2O} is the flow rate of water vapor in the flue gas.

Physically, Equation 10 may be interpreted as follows. The bracketed term on the right hand side is the flow rate of all non-aqueous species in the stack gas, which is determined by subtracting the flow rate of water vapor in the flue gas and the overall gas species absorbtion rate from the total gas phase flue gas flow rate. The fraction term is the ratio of the mole fraction of water vapor in the stack gas to the mole fraction of all other materials in the stack gas. The product of this term with the bracketed flow rate term is the flow rate of water vapor in the stack gas. Subtracting the flow rate of incoming water in the flue gas gives the rate at which water must be vaporized in the scrubber. The flow rate of water in the stack gas (n_{SG,H_2O}) may now be computed as the sum of the water in the flue gas and the water vaporized.

$$n_{SG,H_2O} = n_{FG,H_2O} + \Delta_v n_{SB,H_2O(\ell)} \tag{13}$$

Subroutine GABSO specifies the molar rates of the gas phase species in the stack gas stream with the exception of H_2O. For the species whose fractions absorbed are specified (SO_2, CO_2, NO, and NO_2) the stack gas flow rates will be functions of the quantities not absorbed.

$$n_{SG,jg} = n_{FG,jg}(1.0 - xa_{jg}) \tag{14}$$

for jg = SO_2, CO_2, NO, and NO_2. Two other acid gas species (SO_3 and HCl) are completely absorbed, so that their flow rates in the stack gas will be zero.

$$n_{SG,jg} = 0 \tag{15}$$

for jg = SO_3 and HCl. Nitrogen is not affected in the scrubber, so that the flow rate in the stack gas will be the same as that in the flue gas.

$$n_{SG,N_2} = n_{FG,N_2} \tag{16}$$

As shown in Equations 6 through 8 and the attendant discussion, some oxygen is absorbed in the scrubber to oxidize the NO and NO_2 absorbed to N_2O_5 and that portion of the SO_2

absorbed which is oxidized in the scrubber to SO_3. Based on this requirement, the flow rate of oxygen in the stack gas is calculated by Equation 17.

$$n_{SG,O_2} = n_{FG,O_2} - \Big[{}^3\!/\!_4\, n_{FG,NO}\, xa_{NO} + {}^1\!/\!_4\, n_{FG,NO_2}\, xa_{NO_2} + {}^1\!/\!_2\, n_{FG,SO_2}\, xa_{SO_2}\, xo_S \Big] \qquad (17)$$

This completely specifies the gas phase portion of the stack gas.

The next portion of this routine is concerned with the evaluation of the solid waste stream. The solid waste is composed of all of the materials in the limestone and make-up water streams plus the gases absorbed and particulates removed from the flue gas.

The first step in this section is to determine the particulates remaining in the stack gas. The fractions of the particulates removed from the flue gas (xpart) is specified as an input parameter, so that the molar rates of the solids in the stack gas ($n_{SG,js}$, js = all solid species, gmole/sec) may be specified in terms of the molar rates of the flue gas solids ($n_{FG,js}$) and the fractions not removed.

$$n_{SG,js} = n_{FG,js}(1.0 - \text{xpart}) \qquad (18)$$

for js = all solid species.

The flow rates of calcium and magnesium oxide which pass through the system as solids are now determined. These materials may be present in the flue gas solids and the limestone and may be partially or wholly hydrated and reacted according to the specified input fractions (x/su_{js}, js = CaO and MgO). The flow rates of these solids in the solid waste stream ($n_{SW,js}$, js = CaO and MgO, gmole/sec) will be the sum of these materials from the alkali ($n_{LS,js}$) and flue gas streams multiplied by the fraction not hydrating.

$$n_{SW,js} = \left(n_{LS,js} + \text{xpart } n_{FG,js}\right)\left(1.0 - x/su_{js}\right) \quad (19)$$

for js = CaO and MgO. The flow rate of inerts in the solid waste ($n_{SW,inert}$) is calculated by a similar relation.

$$n_{SW,inert} = n_{LS,inert} + \text{xpart } n_{FG,inert} \quad (20)$$

Subroutine TOSSP1 is now used to calculate the key species for the solids in the flue gas and the limestone streams. This routine sums over the solids for each of the key components. For example, the total calcium oxide flow rate ($n_{S,tCaO(s)}$) will be the sum of the flow rates of all calcium-containing solid species. The reactive key species for these two streams ($n_{S,jt}$) are equal to the solid key species ($n_{S,jt(s)}$) except for CaO and MgO.

$$n_{S,jt} = n_{S,jt(s)} \quad (21)$$

for jt = all key species except CaO and MgO, where S = flue gas and limestone. The total reactive CaO and MgO are determined by subtracting the non-reactive CaO and MgO which pass through the system as solids from the total solid CaO and MgO.

$$n_{S,jt} = n_{S,jt(s)} - n_{S,js} \left(1.0 - x/su_{js}\right) \quad (22)$$

for jt = CaO and MgO while js = CaO and MgO, where S = flue gas and limestone. Subroutine TOSSP1 is also used to calculate the total solid species in the stack gas.

If this is the first pass through this portion of this routine, an estimate of the make-up water flow rate is made. This estimate of the flow rate is based on the assumption that the weight of the solid phase in the solid waste stream will be the same as the weight of the incoming limestone. The make-up water flow rate is set so as to furnish the water to form the necessary weight fraction solids slurry with that weight of material plus the water vaporized in the scrubber.

$$\tilde{L}_{WM} = W_{LS,s} \frac{1.0 - XW_{SW,s}}{XW_{SW,s}} + \Delta_v n_{SB,H_2O(\ell)} MW_{H_2O} \qquad (23)$$

$\tilde{L}_{WM}$ is the flow rate of liquid water in the make-up water stream (g/sec) where the tilde indicates that this quantity is an estimate. $XW_{SW,s}$ is the required weight fraction solids in the solid waste stream, and MW_{H_2O} is the molecular weight of water.

Subroutine TORXSP is not used to evaluate the key species in the make-up water. This routine totals the reacting species in the liquid phase based on the make-up water composition input to SYSTB1 and the flow rate calculated by Equation 23.

The solid waste key species molar rates except for water are now calculated by summing the contributions from the solids in the flue gas and alkali streams, the gases absorbed from the flue gas, and the make-up water, and subtracting the solids which are carried out in the stack gas.

$$n_{SU,jt} = n_{FG,jt} + n_{LS,jt} + \Delta_a n_{FG,jt} + n_{WM,jt} - n_{SG,jt} \quad (24)$$

for jt = all key species.

The weight of nonequilibrium solids in the solid waste is now computed by summing over the molar rates of these solids multiplied by their molecular weights.

$$W_{SW,ns} = \sum_{js} n_{SW,js} MW_{js} \quad (25)$$

for js = all nonequilibrium solid species. The equilibrium program is now called, using the option in which all of the key species rates except water are specified, and the weight fraction solids is specified. The equilibrium program calculates the composition of the solid waste stream, including the total water flow rate as shown in Equation 26.

$$n_{SW,tH_2O} = \Phi\left(T_{AS}, XW_{SW,s}, n_{SW,jt}: \text{ jt = all key species except } H_2O\right) \quad (26)$$

For a limestone simulation, the solid waste pH predicted by the equilibrium program is examined at this point. If this pH is within the tolerance ($\pm pH_{TOL}$) of the desired pH (pH_{MED}) or no limits are set on the solid waste pH (i.e., if $pH_{MED} \leq 0.0$) no action is taken. If the pH is not satisfactory the value of xa_{CO_2} is adjusted by subroutine CNVPH1. Briefly, this routine steps the xa_{CO_2} in the direction to decrease the pH error until the pH median is bracketed. If when the bracket is achieved two or three pH_{SW} - xa_{CO_2} data points are available, linear interpolation is used to predict a new xa_{CO_2}. If more than three points are available arbic interpolation is used. This process is continued until the calculated pH is within the pH tolerance of the desired pH.

Each time the xa_{CO_2} is adjusted, the estimate of the make-up water liquid water flow rate is adjusted. This is done by adding the water vaporized to the total water rate for the solid waste stream predicted by the equilibrium program and multiplying by the molecular weights of water.

$$\tilde{L}_{WM} = \left[n_{SW,tH_2O} + \Delta_v n_{SB,H_2O(\ell)}\right] MW_{H_2O} \tag{27}$$

After this estimate is updated, execution is returned to a point immediately after the evaluation of the stack gas pressure.

If the solid waste pH is satisfactory the make-up water liquid water flow rate is examined. It is required that two subsequent values of this quantity be within 1% of each other. If they are not, the estimate is updated according to Equation 31 and execution is returned to a point immediately after the first estimate of this quantity, Equation 23. When the make-up water flow rate is adequately determined, the equilibrium program is called for the make-up water to determine the detailed composition of that stream.

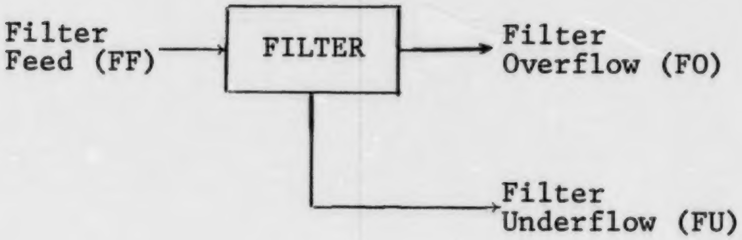
This concludes the calculations for this routine.

Nomenclature:

a	activity
FG	designation for flue gas stream
jg	gaseous species
js	solid species
$\tilde{L}$	liquid flow rate (grams/sec)
L/G	liquid to gas ratio for the system (gpm/1000 acfm)

LISi	label for input stream (i=1,2,3)
LOSi	label for output stream (i=1,2)
LS	designation for alkali stream
$N_{i,j}$	total molar flow rate (i=stream designation j=phase)
$n_{i,j}$	species molar flow rate (i=stream designation j=species)
P	pressure (atm)
SG	designation for stack gas
SW	designation for solid waste stream
U_T	linear gas velocity in the scrubber
vp_{H_2O}	vapor pressure of water
WM	designation for water make-up stream
xa_{jg}	fraction of gaseous species absorbed from the flue gas
xo_s	fraction of the SO_2 absorbed which is oxidized to SO_3 in the scrubber
xpart	fraction of particulates removed
y	mole fraction in the gaseous phase
β_1,β_2,β_3	particulate removal parameters
Δ_v	fraction of water vaporized in the scrubber
$\gamma_1\ \gamma_2\ \gamma_3$	pressure drop parameters

FILTER: SOLID-LIQUID SEPARATOR



Function: This routine simulates a solid-liquid separator on the bases of (1) the filtration efficiency, (2) the weight fraction solids in the filter feed (FF), and (3) complete information about the filter underflow stream (FU). Both the feed and overflow (FO) streams will be calculated.

Input Information: The underflow stream must be completely specified. Two parameters are required: (1) the filtration efficiency (E), and (2) the weight fraction solids in the filter feed ($XW_{FF,s}$).

Output Information: The overflow and feed streams will be completely specified.

Assumptions:

(1) No chemical reaction takes place within the filter.

(2) The filter operates isothermally.

Description: The first step in the calculations for this routine is to set the temperatures (T_{FO} and T_{FF}, °K) and liquid species molalities ($m_{FO,j1}$ and $m_{FF,j1}$, j1 = all liquid species) for the overflow and feed equal to those of the filter underflow (T_{FU} and $m_{FU,jl}$).

$$T_{FO} = T_{FU} \tag{1}$$

$$T_{FF} = T_{FU} \tag{2}$$

$$m_{FO,jl} = m_{FU,jl} \tag{3}$$

$$m_{FF,jl} = m_{FU,jl} \tag{4}$$

for jl = all liquid species.

The efficiency of filtration is defined as the fraction of the solids in the feed which appear in the underflow.. Since there is no chemical reaction within the filter, the compositions of all three streams will be the same, and this relation holds for each solid species as well as for the bulk solid. The molar rates of the solids in the feed ($n_{FF,js}$, js = all solid species) are thus calculated in Equation 5.

$$n_{FF,js} = n_{FU,js}/E \tag{5}$$

for js = all solid species, where $n_{FU,js}$ are the molar rates (gmole/sec) of the solids in the underflow. The molar rates of the solids in the overflow ($n_{FO,js}$) may now be calculated by mass balance difference, as in Equation 6.

$$n_{FO,js} = n_{FF,js} - n_{FU,js} \tag{6}$$

for js = all solid species.

The weight rate of liquid in the filter feed ($W_{FF,l}$) may be found by rearranging the definition of the weight fraction solids ($XW_{FF,s}$), Equation 7, as Equation 8.

$$XW_{FF,s} = \frac{W_{FF,s}}{W_{FF,s} + W_{FF,l}} \tag{7}$$

$$W_{FF,l} = W_{FF,s} \frac{1.0 - XW_{FF,s}}{XW_{FF,s}} \tag{8}$$

The weight of solids in the feed ($W_{FF,s}$, g/sec) used in Equations 7 and 8 may be calculated from the molar rates of the solids determined in Equation 5. The weight fraction solids is specified as an input parameter, so that Equation 8 is a clear statement of the weight of feed liquid.

The flow rate of liquid water in the feed (L_{FF}, g/sec) may be found by ratioing the liquid water rate in the underflow (L_{FU}) by the weight rates of the liquid phases ($W_{FU,l}$ and $W_{FF,l}$) since all three streams have the same phase compositions.

$$L_{FF} = L_{FU}\, W_{FF,l}/W_{FU,l} \tag{9}$$

Equations 8 and 9 may be combined to give Equation 10, which expresses the liquid water flow rate of the filter feed as a function of known quantities.

$$L_{FF} = \frac{L_{FU}}{W_{FU,l}}\, W_{FF,s} \frac{1.0 - XW_{FF,s}}{XW_{FF,s}} \tag{10}$$

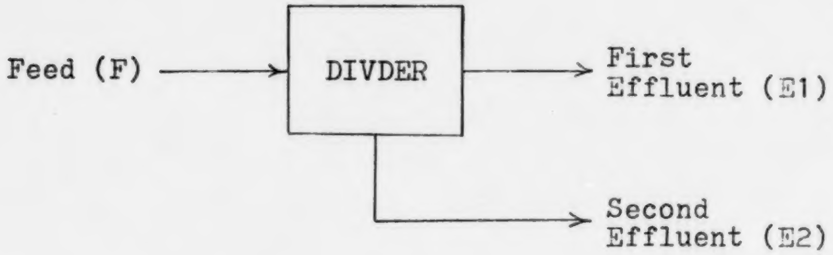
Solving for the liquid water rate in the overflow (L_{FO}) by mass balance difference completes the calculations for this routine.

$$L_{FO} = L_{FF} - L_{FU} \tag{11}$$

Nomenclature:

E	filtration efficiency
FF	designation for the filter feed stream
FO	designation for the filter overflow stream
FU	designation for the filter underflow stream
L	liquid water flow rate
m	molality
n	molar flow rate of a species or component
T	temperature
W	weight flow rate of a stream or phase
XW	weight fraction of a phase

DIVDER: STREAM SPLITTER



Function: This routine simulates a slurry stream splitter on the basis of the flow rate of the first effluent stream (E1) and complete flow and composition information for the second effluent stream (E2).

Input Information: The second effluent stream must be completely specified. The flow rate of the first effluent stream ($\hat{L}_{E1}$, gal/min) must also be specified.

Output Information: Both output streams are completely specified.

Assumptions:

(1) No solid-liquid separation takes place.

(2) No chemical reaction takes place.

Description: The first step in the calculations for this routine is to convert the flow rate of liquid water in the first effluent stream from input units ($\hat{L}_{E1}$, gal/min) to program units (L_{E1}, g/sec).

$$L_{E1} = 453.5924 \cdot 8.2256 \hat{L}_{E1} / 60 \qquad (1)$$

where 8.2245 is the density of liquid water at 55 ^{o}C (131 ^{o}F) in lb/gal. Since the compositions of all three streams are the same and the second effluent stream is completely known, the liquid water flow rate of the feed stream (L_F) may be found by mass balance.

$$L_F = L_{E1} + L_{E2} \qquad (2)$$

where L_{E2} is the liquid water flow rate of the second effluent.

The liquid compositions in the first effluent and feed streams are set equal to the liquid composition in the second effluent stream by setting the molalities in the feed and first effluent ($m_{F,jl}$ and $m_{E1,jl}$) equal to the molalities in the second effluent ($m_{E2,jl}$).

$$m_{F,jl} = m_{E2,jl} \qquad (3)$$

$$m_{E1,jl} = m_{E2,jl} \qquad (4)$$

for jl = all liquid species.

The molar flow rates of the solids and the total reactive species in the first effluent stream ($n_{E1,j}$, j = all solid and key species, gmole/sec) are found by ratioing the corresponding molar rates in the second effluent stream ($n_{E2,j}$) by the liquid water flow rates of the two streams.

$$n_{E1,j} = n_{E2,j} L_{E1} / L_{E2} \qquad (5)$$

for j = all solid and key species. The molar rates of the solids and the total reactive species in the feed stream ($n_{F,j}$) are found by mass balance.

$$n_{F,j} = n_{E1,j} + n_{E2,j} \qquad (6)$$

for j = all solid and key species.

The final step in the calculations is to set the values of a number of intensive quantities for the feed and first effluent stream equal to those of the second effluent stream, as in Equations 7 and 8.

$$q_F = q_{E2} \qquad (7)$$

$$q_{E1} = q_{E2} \qquad (8)$$

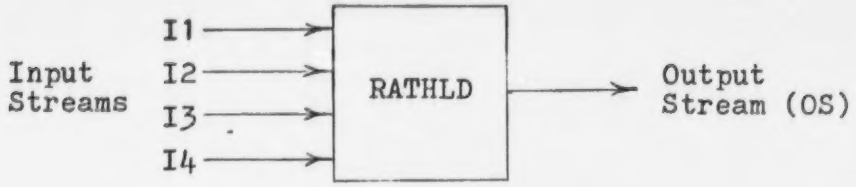
where q is any specified intensive stream quantity. Here, q is (1) temperature, (2) pH, (3) ionic strength, (4) equilibrium partial pressure of CO_2 and SO_2, (5) relative saturation of $CaSO_3 \cdot \frac{1}{2}H_2O$, $CaSO_4 \cdot 2H_2O$, and limestone, and (6) weight fraction solids.

This concludes the calculations for this routine.

Nomenclature:

E1	designation for the first effluent stream
E2	designation for the second effluent stream
F	designation for the feed stream
L	liquid water flow rate
m	molality
n	molar flow rate of a species or component
q	any specified intensive stream quantity
	denotes a quantity not in program units

RATHLD: RATE HOLD TANK



Function: This routine simulates a process hold tank with up to four input streams on the basis of the hold tank volume and complete information about the input streams. A surface-area-dependent solid-liquid mass transfer model is utilized, considering the solids $CaSO_3 \cdot \frac{1}{2}H_2O$, $CaSO_4 \cdot 2H_2O$, and limestone.

Input Information: The input streams must be completely specified. Any unused input streams will be assigned zero values. The tank volume ($\hat{V}$, ft^3) must also be specified.

Output Information: The output stream (OS) will be completely specified.

Description: The first step in the calculations for this routine is to convert the hold tank volume from input units ($\hat{V}$, ft^3) to program units (V, cm^3).

$$V = 28{,}316.85 \ \hat{V} \tag{1}$$

The next step in these calculations is to determine the molar rates of the key species in the effluent stream ($n_{OS,jt}$, jt = all key species, gmole/sec). This calculation is performed by subroutine ADDER, which adds the molar rates of the key species in the input streams ($n_{S,jt}$, for S = I1 through I4).

$$n_{OS,jt} = \sum_{S=I1}^{I4} n_{S,jt} \tag{2}$$

for jt = all key species. Subroutine ADDER also determines the rates at which the solids which are treated by the rate model flow into the hold tank ($n_{I,js}$, js = $CaSO_3 \cdot \frac{1}{2}H_2O$, $CaSO_4 \cdot 2H_2O$, and limestone).

$$n_{I,js} = \sum_{S=I1}^{I4} n_{S,js} \tag{3}$$

for js = $CaSO_3 \cdot \frac{1}{2}H_2O$, $CaSO_4 \cdot 2H_2O$, and limestone.

The rate calculations for these three solids and the equilibrium calculations performed for the liquid phase and other solid species are conducted by subroutine RATE. The rates are calculated based on relative saturation (driving force) as shown in Equation 4.

$$R_{js} = k_{js}\ \Phi(A_{js}, M_{os,js}, T_{os}, V, RS_{js}) \tag{4}$$

where

R_{js} is the precipitation rate (gmoles/sec) of solid js,

k_{js} is the precipitation rate constant (gmoles/sec · cm^2) for solid js, and

Φ is a function of the following variables:

(1) A_{js}, crystallization area per gram of seed

(2) $M_{os,js}$, molality of solid js in the output stream

(3) T_{os}, temperature of the output stream

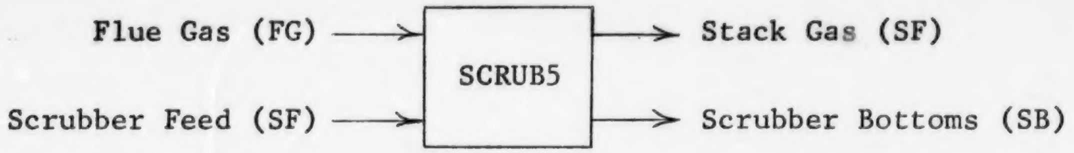
(4) V, volume of the hold tank

(5) RS_{js}, the relative saturation of solid js in the output stream

Nomenclature:

I1 - I4	designation for the input streams
k	precipitation rate constant
m	liquid phase molality
n	molar flow rate of a species or component
OS	designation for the output stream
R	precipitation rate
RS	relative saturation
T	temperature
V	hold tank volume
^	indicates a quantity not in program units

SCRUB5: SCRUBBER



Function: This routine simulates a cocurrent or backmixed gas scrubber on the bases of specified flue gas, stack gas, and scrubber feed. The scrubber bottoms is determined as a function of the other three streams and specified amounts of solid materials reacting. $K_G aV$ and NTU are calculated.

Input Information: The flue gas (FG), stack gas (SG), and scrubber feed (SF) must be specified completely. Several groups of parameters are used.

(1) A mode option is set for either cocurrent or backmixed operation.

(2) A solids option is set to allow or prevent reaction with solid $CaSO_3$, $CaSO_4$, $MgSO_3$, and their hydrates.

(3) The fractions of solid CaO and MgO removed from the flue gas which react in the scrubber must be specified (xh_j).

(4) The fractions of several solids in the scrubber feed and flue gas ($Ca(OH)_2$, $CaCO_3$, $Mg(OH)_2$, $MgCO_3$, $MgCO_3 \cdot 3H_2O$, $MgCO_3 \cdot 5H_2O$, $MgSO_4$, Na_2O,

$CaSO_3$, and limestone) which react in the scrubber must be specified ($xr_{SF,j}$, $sr_{FG,j}$).

(5) The fractions of the lime solids (CaO and MgO) which are used (i.e., hydrated) throughout the system must be specified (xu_j).

Output Information: The scrubber bottoms stream is completely specified. $K_G aV$ and NTU for SO_2 transfer are calculated.

Assumptions:

(1) The scrubber feed, scrubber bottoms, and stack gas streams are at the adiabatic saturation temperature.

(2) Vapor-liquid equilibrium of CO_2 and H_2O occurs.

Description: The calculations for this routine may be grouped into four major units.

(1) The temperatures of the bottoms stream is set.

(2) The reacting contributions to the scrubber bottoms from each input phase are determined.

(3) The composition of the scrubber bottoms is determined.

(4) $K_G aV$ and NTU are calculated.

Each of these steps will now be discussed in detail.

The temperature of the scrubber bottoms is set equal to the temperature of the scrubber feed.

$$T_{SB} = T_{SF} \qquad (1)$$

The reacting materials in the outlet slurry must now be determined. The portions of the scrubber feed materials which react are determined by subroutines XSODIS and DNTSF as functions of $xr_{SF,j}$. The portions of the solids in the flue gas that react are determined by subroutines XSODIS and DNTLA as functions of $xr_{FG,j}$, xh_j, and xu_j. The portions of the flue gas species which are absorbed other than CO_2 are determined in subroutine GABSO. The summation of these contributions is shown as Equation 2.

$$n_{SB,j} = \Delta_r n_{SF,j} + \Delta_r n_{FG,j} + \Delta_r n_{FG,j'} \qquad (2)$$

where j represents all total equilibrium species and j′ is the corresponding gas species.

The CO_2 in the scrubber bottoms is calculated by the equilibrium program as a function of the partial pressure of CO_2 in the stack gas and the moles of other equilibrium species in the scrubber bottoms.

$$n^{*}_{SB,tCO_2(e)} = \Phi[T_{SB}, n_{SB,j}, p_{SG,CO_2}] \qquad (3)$$

for j = all key species except CO_2. The partial pressure is calculated from the total pressure and composition of the stack gas.

$$p_{SG,CO_2} = P_{SG} \cdot n_{SG,CO_2(g)}/N_{SG,g} \qquad (4)$$

where P_{SG} is the total pressure, $n_{SG,CO_2(g)}$ is the molar rate of CO_2, and $N_{SG,g}$ is the total molar gas rate.

The error in the CO_2 mass balance is calculated as the difference between the CO_2 calculated in Equation 2 and Equation 3.

$$\Delta n_{SB,tCO_2(e)} = n_{SB,tCO_2(e)} - n^*_{SB,tCO_2(e)} \qquad (5)$$

This parameter is used in convergence analysis, but is not of further interest here.

The final step in the specification of the scrubber bottoms stream is to add the materials which were withheld from reaction ($\Delta_w n_{SF,j}$ for the scrubber feed and $\Delta_w n_{FG,j}$ for the flue gas) to the materials predicted by the equilibrium program ($n^*_{SB,j}$).

$$n_{SB,j} = n^*_{SB,j} + \Delta_w n_{SF,j} + \Delta_w n_{FG,j} \qquad (6)$$

for j = all solid species.

In order to calculate $K_G aV$ and NTU for SO_2 for the scrubber, it is convenient to first calculate the log mean pressure drop for SO_2 across the scrubber, $\Delta p_{\ell m,SO_2}$. For a cocurrent scrubber, this relation takes the form of Equation 7.

$$\Delta p_{\ell m,SO_2} = \frac{[p_{FG,SO_2} - p^*_{SF,SO_2}] - [p_{SG,SO_2} - p^*_{SB,SO_2}]}{\ln \left[\dfrac{p_{FG,SO_2} - p^*_{SF,SO_2}}{p_{SG,SO_2} - p^*_{SB,SO_2}} \right]} \qquad (7)$$

The $\Delta p_{\ell m,SO_2}$ for a back-mixed contactor is given by Equation 8.

$$\Delta p_{\ell m,SO_2} = \frac{[p_{FG,SO_2} - p^*_{SB,SO_2}] - [p_{SG,SO_2} - p^*_{SB,SO_2}]}{\ln\left[\dfrac{p_{FG,SO_2} - p^*_{SB,SO_2}}{p_{SG,SO_2} - p^*_{SB,SO_2}}\right]} \tag{8}$$

After $\Delta p_{\ell m,SO_2}$ has been calculated by the appropriate equation, NTU and $K_G aV$ may be calculated as in Equations 9 and 10.

$$NTU_{SO_2} = \frac{p_{FG,SO_2} - p_{SG,SO_2}}{\Delta p_{\ell m,SO_2}} \tag{9}$$

$$K_G aV_{SO_2} = \frac{n_{FG,SO_2} - n_{SG,SO_2}}{\Delta p_{\ell m,SO_2}} \tag{10}$$

This completes the calculations for this routine.

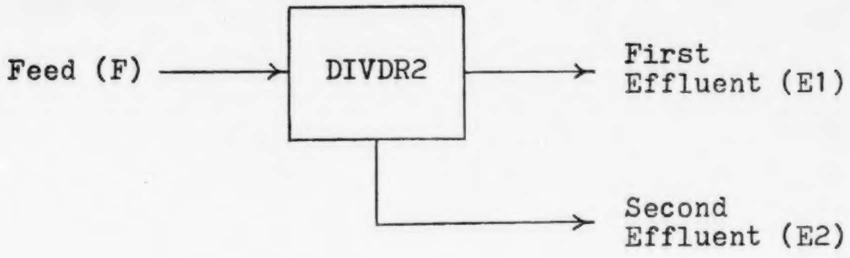
Nomenclature:

*	indicates an equilibrium value
FG	designation for the flue gas stream
$K_G aV$	vapor-liquid mass transfer coefficient
n	molar flow rate of a component
N	molar flow rate of a phase or stream
NTU	number of transfer units
p	partial pressure
P	total pressure
r	indicates reacting material
SB	designation for the scrubber bottoms stream
SF	designation for the scrubber feed stream
SG	designation for the stack gas stream
T	temperature
w	indicates withheld material
xh	fraction of lime solids hydrating
xr	fraction of solids reacting
xu	fraction of limestone utilized

Greek:

Δ	indicates a change or increment
Φ	equilibrium relation

DIVDR2: STREAM SPLITTER



Function: This routine simulates a slurry stream splitter on the basis of the flow rate of the first effluent stream (E1) and complete flow and composition information for the feed stream (F).

Input Information: The feed stream (F) must be completely specified. The flow rate of the first effluent stream ($\hat{L}_{E1}$, gal/min) must also be specified.

Output Information: Both output streams are completely specified.

Assumptions:

(1) No solid-liquid separation takes place.

(2) No chemical reaction takes place.

Description: The first step in the calculations for this routine is to convert the flow rate of liquid water in the first effluent stream from input units ($\hat{L}_{E1}$, gal/min) to program units (L_{E1}, g/sec).

$$L_{E1} = 453.5924 \cdot 8.2256 \hat{L}_{E1} / 60 \qquad (1)$$

where 8.2256 is the density of liquid water at 55 °C (131 °F) in lb/gal. Since the compositions of all three streams are the same and the feed stream is completely known, the liquid water flow rate for the second effluent stream (L_{E2}) may be found by difference.

$$L_{E2} = L_F - L_{E1} \qquad (2)$$

where L_F is the liquid water flow rate of the feed stream.

The liquid compositions in the effluent streams are set equal to the liquid composition in the feed stream by equating the molalities of the species in the effluent streams ($m_{E1,jl}$ and $m_{E2,jl}$) to the molalities in the feed stream ($m_{F,jl}$).

$$m_{E1,jl} = m_{F,jl} \qquad (3)$$

$$m_{E2,jl} = m_{F,jl} \qquad (4)$$

for jl = all liquid species.

The molar flow rates of the solids and the total reactive species in the first effluent stream ($n_{E1,j}$, j = all solid and key species, gmole/sec) are found by ratioing the corresponding molar rates in the feed stream ($n_{F,j}$) by the liquid water flow rates of the two streams.

$$n_{E1,j} = n_{F,j} L_{E1} / L_{F} \tag{5}$$

for j = all solid and key species. The molar flow rates of the solids and the total reactive species in the second effluent stream ($n_{E2,j}$) are found by difference.

$$n_{E2,j} = n_{F,j} - n_{E1,j} \tag{6}$$

for j = all solid and key species.

The final step in the calculations is to set the values of a number of intensive quantities for the output streams equal to those of the input streams, as in Equations 7 and 8.

$$q_{E1} = q_{F} \tag{7}$$

$$q_{E2} = q_{F} \tag{8}$$

where q is any specified intensive stream quantity. Here, q is (1) temperature, (2) pH, (3) ionic strength, (4) equilibrium partial pressure of CO_2 and SO_2, (5) relative saturation of $CaSO_3 \cdot \frac{1}{2}H_2O$, $CaSO_4 \cdot 2H_2O$, and limestone, and (6) weight fraction solids.

This concludes the calculations for this routine.

Nomenclature:

E1	designation for the first effluent stream
E2	designation for the second effluent stream
F	designation for the feed stream
L	liquid water flow rate
m	molality
n	molar flow rate of a species or component
q	any specified intensive stream quantity
^	denotes a quantity not in program units

6.3 APPENDIX III - SIMULATION RESULTS

```
        LG+E SIMULATION NO.5

                  PROCESS DESCRIPTION

EQUIP. NO.     EQUIP. NAME     INPUT STREAMS          OUTPUT STREAMS

    1             ALKINP                                1
    2             FLUGS1                                2
    3             WTRMKP                                4
    4             SYSTB1          2    4    1           3  11
    5             FILTER          7                     8  11
    6             DIVDER         10                     6   7
    7             RATHLD          1    5    8           9
    8             RATHLD          9    4               10
    9             SCRUB5          2    6                3   5
   10             DIVDR2         10                     6   7

               ORDER OF PROCESS CALCULATIONS

      1,2,3(4,5,6(9,7,8,10))*

      RECYCLE LOOP FROM    4  TO  10

      RECYCLE LOOP FROM    9  TO  10
```

```
 07 AUG 75    21:19:14.802
INPUT STREAM DATA

    ALKINP, EQUIPMENT NUMBER  1
         FLOW     7.0200+01 LB/MIN         CASO3    0.0000    WT %
         CAO      0.0000    WT %           MGSO3    0.0000    WT %
         MGO      0.0000    WT %           CASO4    0.0000    WT %
         CACO3    0.0000    WT %           MGSO4    0.0000    WT %
         MGCO3    0.0000    WT %           INERT    0.0000    WT %
         CA(OH)2 9.9940+01 WT %            MG(OH)2 6.0000-02 WT %

    FLUGS1, EQUIPMENT NUMBER  2+  355. DEG F AT  14.700 PSIA
                  FLUE GAS                  ENTRAINED FLY ASH
         FLOW     3.4890+05 ACFM        FLOW     1.2000+01 LB/MIN
         SO2      1.8800-01 MOLE %      CAO        .000    WT %
         CO2      1.1300+01 MOLE %      CASO4      .000    WT %
         NO       0.0000    MOLE %      MGO        .000    WT %
         NO2      0.0000    MOLE %      NA2O       .000    WT %
         O2       6.0000+00 MOLE %      NACL       .000    WT %
         SO3      0.0000    MOLE %      INERT   100.000    WT %
         N2       7.6200+01 MOLE %
         HCL      1.6900-04 MOLE %
         N2O      6.2900+00 MOLE %

    WTRMKP, EQUIPMENT NUMBER  3 +   65. DEG F
         SO3=     1.0000+00 MG/L        CA++     4.6100+01 MG/L
         CO3=     8.8100+01 MG/L        MG++     1.1900+01 MG/L
         SO4=     1.0800+02 MG/L        NA+      3.6400+01 MG/L
         NO3-     1.2500+01 MG/L        CL-      3.6500+01 MG/L
 07 AUG 75    21:19:14.897          LG+E SIMULATION NO.5
SYSTEM AND EQUIPMENT PARAMETERS

   SYSTB1, EQUIPMENT NUMBER  4
        SO2 ABSORBED =   85,00 %        NO  ABSORBED =      ,00 %
        SO2 OXIDIZED =    6,00 %        NO2 ABSORBED =      ,00 %
        LIME SOLIDS HYDRATING IN SYSTEM+  CAO =  100,00 %   MGO =  100,00 %
        PRESSURE DROP CONSTANTS+  GAMMA1 =  7,500-01  GAMMA 2 =  0,000     GAMMA 3 =  0,000
        PARTICULATE REMOVAL CONSTANTS+  BETA 1 =  2,310+00  BETA 2 =  0,000     BETA 3 =  0,000
        INITIAL VALUE OF XA(CO2) =  1,3644-04
        SOLID WASTE PH LIMITED TO  8,00 € 0,5

   FILTER, EQUIPMENT NUMBER  5
        FEED WT FR SOLIDS =  10.00 %   UNDERFLOW WT FR SOLIDS =  40,00 %
        SEPARATION EFFICIENCY = 100,00 %

   DIVDER, EQUIPMENT NUMBER  6
        STREAM NUMBER  6 FLOW =  1.2000+04 GAL/MIN

   SCRUB5, EQUIPMENT NUMBER  9
        BACKMIXED OPERATION      PRECIPITATION NOT ALLOWED
        LIME SOLIDS HYDRATING IN SCRUBBER+  CAO = 100,00 %   MGO = 100,00 %
        FRACTIONS OF SOLIDS AVAILABLE FOR REACTION+
                        CA(OH)2   MG(OH)2    CACO3     MGCO3     CASO3     NA2O      NACL     LIMESTONE
             SLURRY+    100,00 %  100,00 %     ,00 %     ,00 %    1,01 %     ,00 %     ,00 %     ,00 %
             FLUE GAS+     ,00 %     ,00 %     ,00 %     ,00 %     ,00 %     ,00 %     ,00 %     ,00 %

   RATHLD, EQUIPMENT NUMBER  7
        VOLUME =  2,5400+02 CU FT   =  1,9001+03 GAL
        RATE CONSTANTS (GMOLE/SEC CM SQ)+
             CACO3 =  0,0000     CASO4 =  2,0000-06  CASO3 =  1,0000-04  LIMESTONE =  0,0000

   RATHLD, EQUIPMENT NUMBER  8
        VOLUME =  3,0230+04 CU FT   =  2,2614+05 GAL
        RATE CONSTANTS (GMOLE/SEC CM SQ)+
             CACO3 =  0,0000     CASO4 =  2,0000-06  CASO3 =  1,5000-04  LIMESTONE =  0,0000

   DIVDR2, EQUIPMENT NUMBER 10
        STREAM NUMBER  6 FLOW =  1,2000+04 GAL/MIN
```

```
07 AUG 75    21:24:10,674                                  TEMPERATURE    52,123 DEG. C.
                                  INPUT SPECIES (MOLES)

                  H2O  = 6,87714+01                HCL  = 1,65949-02
                  CAO  = 7,16836+00                CO2  = 8,87313-02
                  MGO  = 9,78409-03                N2O3 = 0,00000
                  NA2O = 6,99726-03                N2O5 = 8,90937-04
                                                   SO2  = 6,66174+00
                                                   SO3  = 4,35149-01

                              AQUEOUS SOLUTION EQUILIBRIA

          COMPONENT         MOLALITY          ACTIVITY    ACTIVITY COEFFICIENT

           H+               2,243-07          1,876-07          8,365-01
           H2O                                                  9,991-01
           H2CO3            3,498-04          3,540-04          1,012+00
           HCO3-            1,224-03          9,707-04          7,930-01
           HNO3             1,174-11          1,188-11          1,012+00
           H2SO3            1,996-09          2,020-09          1,012+00
           HSO3-            1,022-04          8,102-05          7,930-01
           HSO4-            2,409-07          1,913-07          7,939-01

           CA++             1,150-02          4,706-03          4,090-01
           CAOH+            5,535-08          4,395-08          7,939-01
           CAHCO3+          1,270-04          1,009-04          7,939-01
           CACO3            3,563-06          3,605-06          1,012+00
           CANO3+           2,327-05          1,847-05          7,939-01
           CASO3            2,874-04          2,908-04          1,012+00
           CASO4            6,273-03          6,347-03          1,012+00

           MG++             3,635-03          1,712-03          4,710-01
           MGOH+            3,830-07          3,041-07          7,939-01
           MGHCO3+          2,214-05          1,758-05          7,939-01
           MGCO3            2,081-06          2,105-06          1,012+00
           MGSO3            3,328-05          3,368-05          1,012+00
           MGSO4            2,723-03          2,755-03          1,012+00

           NA+              8,911-03          7,177-03          8,054-01
           NAOH             6,315-10          6,390-10          1,012+00
           NAHCO3           6,783-06          6,863-06          1,012+00
           NACO3-           3,124-07          2,480-07          7,939-01
           NANO3            2,411-06          2,439-06          1,012+00
           NASO4-           2,559-04          2,032-04          7,930-01

           OH-              4,166-07          3,308-07          7,930-01
           CL-              1,088-02          8,566-03          7,871-01
           CO3--            8,983-07          3,553-07          3,955-01
           NO3-             1,143-03          8,537-04          7,471-01
           SO3--            4,515-05          1,786-05          3,955-01
           SO4--            1,283-02          4,611-03          3,593-01

          COMPONENT         MOLALITY      ACTIVITY PRODUCT  RELATIVE SATURATION

           CA(OH)2(S)       0,000             5,149-16          1,028-10
           CACO3(S)         5.120-02          1,672-09          1,000+00
           CASO3(S)         4,368+00          8,401-08          1,000+00
           CASO4(S)         2,633-01          2,166-05          1,000+00
           MG(OH)2(S)       0,000             1,874-16          2,034-05
           MGCO3(S)         0,000             6,067-10          5,220-05
           MGSO3(S)         0,000             3,050-08          5,530-04

           PCO2 = 1,99072-02  ATM.
           PSO2 = 4,01791-09  ATM.
           PHNO3 = 2,04751-16  ATM.

                           MOLECULAR WATER = 1,52496+00 KGS.

     PH =  6,7267          IONIC STRENGTH = 6,73838-02        RES. E.N. =  -3,873-08

     XS = 40,0000                                             WNES      =   8,328+01

                                      SOLID WASTE
```

```
07 AUG 75    21:25:18,311                                   TEMPERATURE     52,123 DEG. C.
                                  INPUT SPECIES (MOLES)

                  H2O  = 4,10114+04                    HCL  = 8,12761+00
                  CAO  = 2,06339+01                    CO2  = 1,28666+00
                  MGO  = 4,78662+00                    N2O3 = 0,00000
                  NA2O = 3,42131+00                    N2O5 = 4,35604-01
                                                       SO2  = 1,29623+01
                                                       SO3  = 1,00660+01


            SUPERSATURATION ALLOWED

                              AQUEOUS SOLUTION EQUILIBRIA

            COMPONENT           MOLALITY            ACTIVITY     ACTIVITY COEFFICIENT

             H+                 1,946-04            1,611-04            8,282-01
             H2O                                                        9,986-01
             H2CO3              1,733-03            1,761-03            1,016+00
             HCO3-              7,285-06            5,623-06            7,718-01
             HNO3               9,588-09            9,745-09            1,016+00
             H2SO3              2,792-04            2,837-04            1,016+00
             HSO3-              1,717-02            1,325-02            7,718-01
             HSO4-              1,888-04            1,463-04            7,750-01

             CA++               1,932-02            7,179-03            3,716-01
             CAOH+              1,007-10            7,803-11            7,750-01
             CAHCO3+            1,150-06            8,914-07            7,750-01
             CACO3              3,650-11            3,710-11            1,016+00
             CANO3+             3,474-05            2,692-05            7,750-01
             CASO3              8,313-05            8,448-05            1,016+00
             CASO4              8,488-03            8,626-03            1,016+00

             MG++               3,994-03            1,757-03            4,398-01
             MGOH+              4,685-10            3,631-10            7,750-01
             MGHCO3+            1,348-07            1,045-07            7,750-01
             MGCO3              1,434-11            1,457-11            1,016+00
             MGSO3              6,474-06            6,579-06            1,016+00
             MGSO4              2,478-03            2,518-03            1,016+00

             NA+                9,028-03            7,110-03            7,876-01
             NAOH               7,249-13            7,367-13            1,016+00
             NAHCO3             3,876-08            3,939-08            1,016+00
             NACO3-             2,139-12            1,657-12            7,750-01
             NANO3              2,272-06            2,309-06            1,016+00
             NASO4-             2,314-04            1,793-04            7,750-01

             OH-                4,967-10            3,850-10            7,750-01
             CL-                1,100-02            8,412-03            7,646-01
             CO3--              6,753-12            2,397-12            3,549-01
             NO3-               1,142-03            8,156-04            7,141-01
             SO3--              9,583-06            3,401-06            3,549-01
             SO4--              1,307-02            4,108-03            3,144-01

            COMPONENT           MOLALITY        ACTIVITY PRODUCT  RELATIVE SATURATION

             CA(OH)2(S)         0,000               1,064-21            3,364-10
             CACO3(S)           0,000               1,721-14            1,029-05
             CASO3(S)           0,000               2,440-08            2,984-01
             CASO4(S)           0,000               2,941-05            1,358+00
             MG(OH)2(S)         0,000               2,604-22            2,827-11
             MGCO3(S)           0,000               4,193-15            3,687-10
             MGSO3(S)           0,000               5,950-09            1,079-04

             PCO2 = 9,90889-02  ATM.
             PSO2 = 5,64598-04  ATM.
             PHNO3 = 1,68010-13  ATM.

                               MOLECULAR WATER = 7,38790+02 KGS.

        PH =  3,7928          IONIC STRENGTH = 9,22834-02        RES. E.N. =   5,169-07

                                  SCRUBS, EQUIPMENT NUMBER  9
                                        BOTTOMS STREAM

   DELTA PCO2 =  7,24557-03       DESIRED PCO2 =  9,83761-02

   PCO2 ERROR =  7,36517+00     XA(CO2) =  1,35201-04
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                         RATHLD, EQUIPMENT NUMBER  7

07 AUG 75    21:25:20,345                            TEMPERATURE    52,123 DEG, C,
                           INPUT SPECIES (MOLES)

              H2O  = 4,17735+04            HCL  = 8,21058+00
              CAO  = 6,03214+02            CO2  = 7,74448+00
              MGO  = 4,84088+00            N2O3 = 0,00000
              NA2O = 3,45613+00            N2O5 = 4,40034-01
                                           SO2  = 5,50229+02
                                           SO3  = 4,98088+01

                        AQUEOUS SOLUTION EQUILIBRIA

      COMPONENT         MOLALITY         ACTIVITY    ACTIVITY COEFFICIENT

       H+               1,048-07         8,719-08          8,320-01
       H2O                                                 9,988-01
       H2CO3            6,106-05         6,191-05          1,014+00
       HCO3-            4,669-04         3,653-04          7,825-01
       HNO3             5,349-12         5,423-12          1,014+00
       H2SO3            1,370-08         1,389-08          1,014+00
       HSO3-            1,532-03         1,199-03          7,825-01
       HSO4-            1,152-07         9,039-08          7,844-01

       CA++             1,490-02         5,810-03          3,900-01
       CAOH+            1,488-07         1,167-07          7,844-01
       CAHCO3+          5,975-05         4,687-05          7,844-01
       CACO3            3,555-06         3,605-06          1,014+00
       CANO3+           2,857-05         2,241-05          7,844-01
       CASO3            1,127-02         1,143-02          1,014+00
       CASO4            7,860-03         7,969-03          1,014+00

       MG++             3,212-03         1,462-03          4,551-01
       MGOH+            7,119-07         5,584-07          7,844-01
       MGHCO3+          7,200-06         5,648-06          7,844-01
       MGCO3            1,436-06         1,456-06          1,014+00
       MGSO3            9,028-04         9,153-04          1,014+00
       MGSO4            2,359-03         2,392-03          1,014+00

       NA+              8,989-03         7,160-03          7,965-01
       NAOH             1,352-09         1,371-09          1,014+00
       NAHCO3           2,542-06         2,577-06          1,014+00
       NACO3-           2,555-07         2,004-07          7,844-01
       NANO3            2,358-06         2,391-06          1,014+00
       NASO4-           2,628-04         2,061-04          7,844-01

       OH-              9,071-07         7,116-07          7,844-01
       CL-              1,100-02         8,533-03          7,760-01
       CO3--            7,674-07         2,878-07          3,750-01
       NO3-             1,148-03         8,389-04          7,310-01
       SO3--            1,516-03         5,686-04          3,750-01
       SO4--            1,393-02         4,689-03          3,366-01

      COMPONENT         MOLALITY     ACTIVITY PRODUCT  RELATIVE SATURATION

       CA(OH)2(S)       0,000            2,942-15          9,381-10
       CACO3(S)         9,769-03         1,672-09          9,999-01
       CASO3(S)         7,217-01         3,301-06          3,930+01
       CASO4(S)         4,229-02         2,718-05          1,255+00
       MG(OH)2(S)       0,000            7,401-16          8,037-05
       MGCO3(S)         0,000            4,191-10          3,606-05
       MGSO3(S)         0,000            8,281-07          1,582-02

       PCO2 = 3,48302-03  ATM,
       PSO2 = 2,76335-08  ATM,
       PHNO3 = 9,35048-17  ATM,

                    MOLECULAR WATER = 7,46658+02 KGS,

PH =  7,0595         IONIC STRENGTH = 7,88631-02       RES, E,N, =   9,499-08

                              RATE PARAMETERS

      COMPONENT         INPUT RATE       RATE CONSTANT      PRECIPITATION
                         GMOLE/SEC        KG H2O/SEC      RATE, GMOLE/SEC

       CASO3(S)         5,373+02         5,777-02          1,597+00
       CASO4(S)         3,157+01         3,971-01          4,284-03
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                              RATHLD, EQUIPMENT NUMBER  0

07 AUG 75    21:25:26,938                                     TEMPERATURE    52,123 DEG, C,
                                  INPUT SPECIES (MOLES)

                     H2O  = 4,22634+04            HCL  = 8,21968+00
                     CAO  = 6,03223+02            CO2  = 7,75744+00
                     MGO  = 4,84498+00            N2O3 = 0,00000
                     NA2O = 3,46295+00            N2O5 = 4,40903-01
                                                  SO2  = 5,50229+02
                                                  SO3  = 4,98179+01

                              AQUEOUS SOLUTION EQUILIBRIA

              COMPONENT          MOLALITY          ACTIVITY     ACTIVITY COEFFICIENT

               H+                2,251-07          1,879-07          8,348-01
               H2O                                                   9,990-01
               H2CO3             3,262-04          3,303-04          1,013+00
               HCO3-             1,146-03          9,044-04          7,892-01
               HNO3              1,163-11          1,178-11          1,013+00
               H2SO3             4,477-09          4,534-09          1,013+00
               HSO3-             2,300-04          1,816-04          7,892-01
               HSO4-             2,535-07          2,004-07          7,905-01

               CA++              1,258-02          5,059-03          4,020-01
               CAOH+             5,967-08          4,716-08          7,905-01
               CAHCO3+           1,278-04          1,010-04          7,905-01
               CACO3             3,560-06          3,605-06          1,013+00
               CANO3+            2,487-05          1,966-05          7,905-01
               CASO3             6,908-04          6,995-04          1,013+00
               CASO4             7,050-03          7,138-03          1,013+00

               MG++              3,564-03          1,658-03          4,651-01
               MGOH+             3,718-07          2,939-07          7,905-01
               MGHCO3+           2,006-05          1,586-05          7,905-01
               MGCO3             1,872-06          1,896-06          1,013+00
               MGSO3             7,204-05          7,295-05          1,013+00
               MGSO4             2,756-03          2,790-03          1,013+00

               NA+               8,892-03          7,133-03          8,021-01
               NAOH              6,261-10          6,340-10          1,013+00
               NAHCO3            6,276-06          6,355-06          1,013+00
               NACO3-            2,901-07          2,293-07          7,905-01
               NANO3             2,370-06          2,400-06          1,013+00
               NASO4-            2,672-04          2,112-04          7,905-01

               OH-               4,178-07          3,302-07          7,905-01
               CL-               1,088-02          8,521-03          7,831-01
               CO3--             8,519-07          3,305-07          3,880-01
               NO3-              1,140-03          8,451-04          7,413-01
               SO3--             1,030-04          3,996-05          3,880-01
               SO4--             1,374-02          4,824-03          3,510-01

              COMPONENT          MOLALITY      ACTIVITY PRODUCT  RELATIVE SATURATION

               CA(OH)2(S)        0,000             5,517-16          1,744-10
               CACO3(S)          8,637-03          1,672-09          1,000+00
               CASO3(S)          7,273-01          2,020-07          2,405+00
               CASO4(S)          4,213-02          2,436-05          1,125+00
               MG(OH)2(S)        0,000             1,808-16          1,963-05
               MGCO3(S)          0,000             5,463-10          4,700-05
               MGSO3(S)          0,000             6,605-06          1,198-03

               PCO2 = 1,85789-02  ATM,
               PSO2 = 9,01909-09  ATM,
               PHNO3 = 2,03020-16  ATM,

                             MOLECULAR WATER = 7,55382+02 KGS,

         PH =  6,7260          IONIC STRENGTH = 7,13575-02        RES, E,N, =  -2,400-08

                                     RATE PARAMETERS

              COMPONENT          INPUT RATE        RATE CONSTANT        PRECIPITATION
                                 GMOLE/SEC          KG H2O/SEC       RATE, GMOLE/SEC

               CASO3(S)          5,389+02           1,031+01           1,054+01
               CASO4(S)          3,150+01           4,726+01           2,480-01
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07 AUG 75    21:25:31.221                         LG+E SIMULATION NO.5
 STREAM NUMBER                         1           2           3           4           5           6
 TOTAL STREAM
    FLOW RATE (G/SEC)               530.71      .13090+06   .13771+06   8830.1      .82545+06   .83225+06
              (G-MOLE/SEC)          7.1637      4436.0      4834.9      490.02      41747.      42139.
    TEMPERATURE (DEG. K)            325.28      452.60      325.28      325.28      325.28      325.28
    PRESSURE  (ATM)                 .00000      1.0003      .94924      .00000      .00000      .00000
    ENTHALPY  (CAL/SEC)            -.16835+07  -.59063+08  -.85968+08  -.33240+08  -.30000+10  -.30268+10
    SOLIDS    (WT %)                100.000     .69306-01   .53990-02   .00000      9.9760      9.9667
              (MG/CU M)             .00000      550.95      54.688      .00000      .00000      .00000
 GAS PHASE
    FLOW RATE (G/SEC)               .00000      .13081+06   .13770+06   .00000      .00000      .00000
              (G-MOLE/SEC)          .00000      4434.8      4834.8      .00000      .00000      .00000
              (L/SEC AT T)          .00000      .16466+06   .13595+06   .00000      .00000      .00000
              (L/SEC, STP)          .00000      99403.      .10837+06   .00000      .00000      .00000
    ENTHALPY  (CAL/SEC)             .00000     -.58776+08  -.85945+08   .00000      .00000      .00000
    DENSITY   (G/L AT T)            .00000      .79441      1.0129      .00000      .00000      .00000
    COMP      (G-MOLES/SEC)
      SO2                           .00000      8.3375      1.2506      .00000      .00000      .00000
      CO2                           .00000      501.13      501.07      .00000      .00000      .00000
      NO                            .00000      .00000      .00000      .00000      .00000      .00000
      NO2                           .00000      .00000      .00000      .00000      .00000      .00000
      O2                            .00000      266.09      265.88      .00000      .00000      .00000
      SO3                           .00000      .00000      .00000      .00000      .00000      .00000
      N2                            .00000      3379.3      3379.3      .00000      .00000      .00000
      HCL                           .00000      .74948-02   .00000      .00000      .00000      .00000
      H2O                           .00000      278.95      687.30      .00000      .00000      .00000
      CH4                           .00000      .00000      .00000      .00000      .00000      .00000
    COMP      (MOLE %)
      SO2                           .00000      .18800      .25867-01   .00000      .00000      .00000
      CO2                           .00000      11.300      10.364      .00000      .00000      .00000
      NO                            .00000      .00000      .00000      .00000      .00000      .00000
      NO2                           .00000      .00000      .00000      .00000      .00000      .00000
      O2                            .00000      6.0000      5.4992      .00000      .00000      .00000
      SO3                           .00000      .00000      .00000      .00000      .00000      .00000
      N2                            .00000      76.200      69.896      .00000      .00000      .00000
      HCL                           .00000      .16900-03   .00000      .00000      .00000      .00000
      H2O                           .00000      6.2900      14.216      .00000      .00000      .00000
      CH4                           .00000      .00000      .00000      .00000      .00000      .00000
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07 AUG 75     21:25:31.443                         LG+E SIMULATION NO.5
 STREAM NUMBER                        1            2            3            4            5            6
 SLURRY STREAM
    FLOW RATE (G/SEC)             .00000       .00000       .00000       .00000      .82545+06    .83225+06
              (G-MOLE/SEC)        .00000       .00000       .00000       .00000      41747.       42139.
              (L/SEC AT T)        .00000       .00000       .00000       .00000      748.46       755.99
    ENTHALPY  (CAL/SEC)           .00000       .00000       .00000       .00000     -.30000+10   -.30268+10
    BULK DENSITY(G/ML AT T)       .00000       .00000       .00000       .00000      1102.9       1100.9
    KEY COMP RATE(GMOLES/SEC)
      SO2                         .00000       .00000       .00000       .00000      550.23       543.56
      CO2                         .00000       .00000       .00000       .00000      7.7312       7.6634
      SO3                         .00000       .00000       .00000       .00000      49.640       49.214
      N2O5                        .00000       .00000       .00000       .00000      .43558       .43554
      CAO                         .00000       .00000       .00000       .00000      595.92       595.91
      MGO                         .00000       .00000       .00000       .00000      4.7865       4.7863
      NA2O                        .00000       .00000       .00000       .00000      3.4211       3.4208
      HCL                         .00000       .00000       .00000       .00000      8.1276       8.1201
      H2O                         .00000       .00000       .00000       .00000      41343.       41751.
 LIQUID PHASE
    FLOW RATE (G/SEC)             .00000       .00000       .00000      8830.1       .74310+06    .74930+06
              (G-MOLES/SEC)       .00000       .00000       .00000      490.02       41081.       41469.
    ENTHALPY  (CAL/SEC)           .00000       .00000       .00000     -.33240+08   -.27915+10   -.28169+10
    DENSITY   (G/ML AT T)         .00000       .00000       .00000      .98742       .99284       .99115
    KEY COMP RATE(G-MOLES/SEC)
      SO2                         .00000       .00000       .00000      .11042-03    12.962       .81779
      CO2                         .00000       .00000       .00000      .12979-01    1.2866       1.2185
      SO3                         .00000       .00000       .00000      .99384-02    18.065       17.773
      N2O5                        .00000       .00000       .00000      .89089-03    .43558       .43554
      CAO                         .00000       .00000       .00000      .10168-01    20.633       15.283
      MGO                         .00000       .00000       .00000      .43256-02    4.7865       4.7863
      NA2O                        .00000       .00000       .00000      .69969-02    3.4211       3.4208
      HCL                         .00000       .00000       .00000      .91000-02    8.1276       8.1201
      H2O                         .00000       .00000       .00000      489.96       41011.       41417.
    KEY COMP CONC(G-MOLES/KG H2O)
      SO2                         .00000       .00000       .00000      .12509-04    .17545-01    .10959-02
      CO2                         .00000       .00000       .00000      .14703-02    .17415-02    .16329-02
      SO3                         .00000       .00000       .00000      .11259-02    .24453-01    .23817-01
      N2O5                        .00000       .00000       .00000      .10093-03    .58958-03    .58366-03
      CAO                         .00000       .00000       .00000      .11519-02    .27929-01    .20481-01
      MGO                         .00000       .00000       .00000      .49003-03    .64788-02    .64141-02
      NA2O                        .00000       .00000       .00000      .79266-03    .46307-02    .45842-02
      HCL                         .00000       .00000       .00000      .10309-02    .11001-01    .10881-01
    PH                            .00000       .00000       .00000      7.3046       3.7928       6.7260
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07 AUG 75    21:25:31.684                               LG+E SIMULATION NO. 5
 STREAM NUMBER                        1           2           3           4           5           6
 LIQUID PHASE(CONT)
    LIQ. H2O RATE (KG/SEC)        .00000      .00000      .00000      8.8271      738.79      746.23
                 (LITERS/SEC)     .00000      .00000      .00000      8.9395      744.12      752.89
    CONC          (G-MOLES/KG H2O)
      H+                          .00000      .00000      .00000      .53956-07   .19457-03   .22511-06
      OH-                         .00000      .00000      .00000      .13731-05   .49673-09   .41778-06
      HSO3-                       .00000      .00000      .00000      .29989-05   .17166-01   .23004-03
      SO3=                        .00000      .00000      .00000      .32861-05   .95827-05   .10299-03
      SO4=                        .00000      .00000      .00000      .92525-03   .13067-01   .13744-01
      HCO3-                       .00000      .00000      .00000      .13212-02   .72854-05   .11459-02
      CO3=                        .00000      .00000      .00000      .24039-05   .00000      .85189-06
      NO3-                        .00000      .00000      .00000      .20111-03   .11422-02   .11401-02
      HSO4-                       .00000      .00000      .00000      .76211-08   .18881-03   .25353-06
      H2SO3(L)                    .00000      .00000      .00000      .18022-10   .27915-03   .44774-08
      H2CO3(L)                    .00000      .00000      .00000      .11613-03   .17329-02   .32623-03
      CA++                        .00000      .00000      .00000      .99454-03   .19322-01   .12584-01
      CAOH+                       .00000      .00000      .00000      .26884-07   .10068-09   .59666-07
      CASO3(L)                    .00000      .00000      .00000      .54740-05   .83130-04   .69082-03
      CACO3(L)                    .00000      .00000      .00000      .24951-05   .36502-10   .35604-05
      CAHCO3+                     .00000      .00000      .00000      .20252-04   .11502-05   .12781-03
      CASO4(L)                    .00000      .00000      .00000      .12848-03   .84880-02   .70496-02
      CANO3+                      .00000      .00000      .00000      .63703-06   .34739-04   .24869-04
      MG++                        .00000      .00000      .00000      .41716-03   .39944-02   .35639-02
      MGOH+                       .00000      .00000      .00000      .22014-06   .46847-09   .37181-06
      MGSO3(L)                    .00000      .00000      .00000      .75013-06   .64736-05   .72042-04
      MGHCO3+                     .00000      .00000      .00000      .41764-05   .13479-06   .20058-04
      MGSO4(L)                    .00000      .00000      .00000      .66002-04   .24778-02   .27559-02
      MGCO3(L)                    .00000      .00000      .00000      .17243-05   .14335-10   .18725-05
      NA+                         .00000      .00000      .00000      .15771-02   .90278-02   .88922-02
      NAOH(L)                     .00000      .00000      .00000      .48581-09   .72494-12   .62610-09
      NACO3-                      .00000      .00000      .00000      .25701-06   .21385-11   .29008-06
      NAHCO3(L)                   .00000      .00000      .00000      .17126-05   .38759-07   .62764-05
      NASO4-                      .00000      .00000      .00000      .61576-05   .23137-03   .26723-03
      NANO3(L)                    .00000      .00000      .00000      .10454-06   .22718-05   .23700-05
      CL-                         .00000      .00000      .00000      .10309-02   .11001-01   .10881-01
    IONIC STRENGTH                .00000      .00000      .00000      .67685-02   .92283-01   .71357-01
    DISSOLVED SOLIDS (WPPM)       .00000      .00000      .00000      342.54      5799.3      4104.9
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07 AUG 75    21:25:31.930                         LG+E SIMULATION NO.5
 STREAM NUMBER                       1           2           3           4           5           6
 SOLID PHASE
   FLOW RATE (G/SEC)              530.71      90.718      7.4348      .00000      82347.      82948.
             (G-MOLE/SEC)         7.1637      1.1968      .98084-01   .00000      665.99      670.23
   ENTHALPY  (CAL/SEC)          -.16835+07  -.28747+06  -23754.      .00000    -.20851+09  -.20991+09
   SPECIFIC GRAVITY               2.2401      3.2800      3.2800      .00000      2.5465      2.5453
   KEY COMP RATE(GMOLES/SEC)
     SO2                          .00000      .00000      .00000      .00000      537.26      542.74
     CO2                          .00000      .00000      .00000      .00000      6.4446      6.4450
     SO3                          .00000      .00000      .00000      .00000      31.575      31.442
     N2O5                         .00000      .00000      .00000      .00000      .00000      .00000
     CAO                          7.1582      .00000      .00000      .00000      575.28      580.63
     MGO                          .54585-02   .00000      .00000      .00000      .00000      .00000
     NA2O                         .00000      .00000      .00000      .00000      .00000      .00000
     HCL                          .00000      .00000      .00000      .00000      .00000      .00000
     H2O                          7.1637      .00000      .00000      .00000      331.78      334.25
   COMP RATE    (G MOLES/SEC)
     CAO                          .00000      .00000      .00000      .00000      .00000      .00000
     CA(OH)2                      7.1582      .00000      .00000      .00000      .00000      .00000
     CACO3                        .00000      .00000      .00000      .00000      6.4446      6.4450
     CASO3                        .00000      .00000      .00000      .00000      .00000      .00000
     CASO3*1/2H2O                 .00000      .00000      .00000      .00000      537.26      542.74
     CASO4                        .00000      .00000      .00000      .00000      .00000      .00000
     CASO4*2H2O                   .00000      .00000      .00000      .00000      31.575      31.442
     MGO                          .00000      .00000      .00000      .00000      .00000      .00000
     MG(OH)2                      .54585-02   .00000      .00000      .00000      .00000      .00000
     MGCO3                        .00000      .00000      .00000      .00000      .00000      .00000
     MGCO3*3H2O                   .00000      .00000      .00000      .00000      .00000      .00000
     MGCO3*5H2O                   .00000      .00000      .00000      .00000      .00000      .00000
     MGSO3                        .00000      .00000      .00000      .00000      .00000      .00000
     MGSO3*3H2O                   .00000      .00000      .00000      .00000      .00000      .00000
     MGSO3*6H2O                   .00000      .00000      .00000      .00000      .00000      .00000
     MGSO4                        .00000      .00000      .00000      .00000      .00000      .00000
     MGCL2                        .00000      .00000      .00000      .00000      .00000      .00000
     NA2O                         .00000      .00000      .00000      .00000      .00000      .00000
     NACL                         .00000      .00000      .00000      .00000      .00000      .00000
     INERTS                       .00000      1.1968      .98084-01   .00000      90.702      89.603
     LIMESTONE                    .00000      .00000      .00000      .00000      .00000      .00000
   XLS                            .00000      .00000      .00000      .00000      .00000      .00000
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07 AUG 75    21:25:32.155                      LG+E SIMULATION NO.5
 STREAM NUMBER                   1           2           3            4            5            6
 ELEMENTS(G-ATOMS/SEC
 TOTAL STREAM
      H                     14.327      557.91      1374.6       979.93       82694.       83511.
      C                     .00000      501.13      501.07       .12979-01    7.7312       7.6634
      N                     .00000      6758.7      6758.7       .17818-02    .87116       .87108
      O                     14.327      1830.1      2223.7       490.04       43214.       43608.
      NA                    .00000      .00000      .00000       .13994-01    6.8423       6.8416
      MG                    .54585-02   .00000      .00000       .43256-02    4.7865       4.7863
      S                     .00000      8.3375      1.2506       .10049-01    599.87       592.77
      CL                    .00000      .74948-02   .00000       .91000-02    8.1276       8.1201
      CA                    7.1582      .00000      .00000       .10168-01    595.92       595.91
 GAS PHASE
      H                     .00000      557.91      1374.6       .00000       .00000       .00000
      C                     .00000      501.13      501.07       .00000       .00000       .00000
      N                     .00000      6758.7      6758.7       .00000       .00000       .00000
      O                     .00000      1830.1      2223.7       .00000       .00000       .00000
      S                     .00000      8.3375      1.2506       .00000       .00000       .00000
      CL                    .00000      .74948-02   .00000       .00000       .00000       .00000
 LIQUID PHASE
      H                     .00000      .00000      .00000       979.93       82031.       82842.
      C                     .00000      .00000      .00000       .12979-01    1.2866       1.2185
      N                     .00000      .00000      .00000       .17818-02    .87116       .87108
      O                     .00000      .00000      .00000       490.04       41125.       41500.
      NA                    .00000      .00000      .00000       .13994-01    6.8423       6.8416
      MG                    .00000      .00000      .00000       .43256-02    4.7865       4.7863
      S                     .00000      .00000      .00000       .10049-01    31.027       18.591
      CL                    .00000      .00000      .00000       .91000-02    8.1276       8.1201
      CA                    .00000      .00000      .00000       .10168-01    20.633       15.283
 SOLID PHASE
      H                     14.327      .00000      .00000       .00000       663.56       668.51
      C                     .00000      .00000      .00000       .00000       6.4446       6.4450
      N                     .00000      .00000      .00000       .00000       .00000       .00000
      O                     14.327      .00000      .00000       .00000       2089.2       2107.6
      NA                    .00000      .00000      .00000       .00000       .00000       .00000
      MG                    .54585-02   .00000      .00000       .00000       .00000       .00000
      S                     .00000      .00000      .00000       .00000       568.84       574.18
      CL                    .00000      .00000      .00000       .00000       .00000       .00000
      CA                    7.1582      .00000      .00000       .00000       575.28       580.63
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07 AUG 75    21:25:32.392                         LG+E SIMULATION NO.5
 STREAM NUMBER                  1            2            3            4            5            6
 TOTAL STREAM
   FLOW RATE (LBS/HR)        4212.1       .10389+07    .10929+07    70082.       .65513+07    .66053+07
             (LB-MOLES/HR)   56.855       35207.       38373.       3889.1       .33133+06    .33444+06
   TEMPERATURE(DEG F)        125.82       355.00       125.82       125.82       125.82       125.82
   PRESSURE  (PSIA)          .00000       14.700       13.950       .00000       .00000       .00000
   ENTHALPY  (BTU/HR)       -.24051+08   -.84378+09   -.12281+10   -.47487+09   -.42857+11   -.43241+11
   SOLIDS    (WT % )         100.000      .69306-01    .53990-02    .00000       9.9760       9.9667
             (GR/ACF, WET)   .00000       .24076       .23898-01    .00000       .00000       .00000
 GAS PHASE
   FLOW RATE (LBS/HR)        .00000       .10382+07    .10929+07    .00000       .00000       .00000
             (LB-MOLES/HR)   .00000       35198.       38372.       .00000       .00000       .00000
             (ACFM)          .00000       .34890+06    .28807+06    .00000       .00000       .00000
             (SCFM)          .00000       .21051+06    .25484+06    .00000       .00000       .00000
   ENTHALPY  (BTU/HR)        .00000      -.83967+09   -.12278+10    .00000       .00000       .00000
   DENSITY   (LBS/FT3)       .00000       .66293-02    .84524-02    .00000       .00000       .00000
   COMP      (MOLE %)
     SO2                     .00000       .18800       .25867-01    .00000       .00000       .00000
     CO2                     .00000       11.300       10.364       .00000       .00000       .00000
     NO                      .00000       .00000       .00000       .00000       .00000       .00000
     NO2                     .00000       .00000       .00000       .00000       .00000       .00000
     O2                      .00000       6.0000       5.4992       .00000       .00000       .00000
     SO3                     .00000       .00000       .00000       .00000       .00000       .00000
     N2                      .00000       76.200       69.896       .00000       .00000       .00000
     HCL                     .00000       .16900-03    .00000       .00000       .00000       .00000
     H2O                     .00000       6.2900       14.216       .00000       .00000       .00000
     CH4                     .00000       .00000       .00000       .00000       .00000       .00000
 LIQUID PHASE
   FLOW RATE (LBS/HR)        .00000       .00000       .00000       70082.       .58977+07    .59469+07
             (LB-MOLES/HR)   .00000       .00000       .00000       3889.1       .32605+06    .32912+06
   ENTHALPY  (BTU/HR)        .00000       .00000       .00000      -.47487+09   -.39879+11   -.40242+11
   DENSITY   (LBS/GAL)       .00000       .00000       .00000       8.2401       8.2853       8.2712
   KEY COMP CONC(MOLALITY)
     SO2                     .00000       .00000       .00000       .12509-04    .17545-01    .10959-02
     CO2                     .00000       .00000       .00000       .14703-02    .17415-02    .16329-02
     SO3                     .00000       .00000       .00000       .11259-02    .24453-01    .23817-01
     N2O5                    .00000       .00000       .00000       .10093-03    .58958-03    .58366-03
     CAO                     .00000       .00000       .00000       .11519-02    .27929-01    .20481-01
     MGO                     .00000       .00000       .00000       .49003-03    .64788-02    .64141-02
     NA2O                    .00000       .00000       .00000       .79266-03    .46307-02    .45842-02
     HCL                     .00000       .00000       .00000       .10309-02    .11001-01    .10881-01
   PH                        .00000       .00000       .00000       7.3046       3.7928       6.7260
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07 AUG 75    21:25:32.633                        LG+E SIMULATION NO.5
 STREAM NUMBER                      1          2          3          4           5           6
 LIQUID PHASE(CONT)
    LIQ H2O RATE ( LBS/HR)      .00000     .00000     .00000     70058.      .58635+07   .59225+07
                 (GAL/MIN)      .00000     .00000     .00000     141.75      11864.      11983.
    CONC (MOLALITY)
      H+                        .00000     .00000     .00000     .53956-07   .19457-03   .22511-06
      OH-                       .00000     .00000     .00000     .13731-05   .49673-09   .41778-06
      HSO3-                     .00000     .00000     .00000     .29989-05   .17166-01   .23004-03
      SO3=                      .00000     .00000     .00000     .32861-05   .95827-05   .10299-03
      SO4=                      .00000     .00000     .00000     .92525-03   .13067-01   .13744-01
      HCO3-                     .00000     .00000     .00000     .13212-02   .72854-05   .11459-02
      CO3=                      .00000     .00000     .00000     .24039-05   .00000      .85189-06
      NO3-                      .00000     .00000     .00000     .20111-03   .11422-02   .11401-02
      HSO4-                     .00000     .00000     .00000     .76211-08   .18881-03   .25353-06
      H2SO3(L)                  .00000     .00000     .00000     .18022-10   .27915-03   .44774-08
      H2CO3(L)                  .00000     .00000     .00000     .11613-03   .17329-02   .32623-03
      CA++                      .00000     .00000     .00000     .99454-03   .19322-01   .12584-01
      CAOH+                     .00000     .00000     .00000     .26884-07   .10068-09   .59666-07
      CASO3(L)                  .00000     .00000     .00000     .54740-05   .83130-04   .69082-03
      CACO3(L)                  .00000     .00000     .00000     .24951-05   .36502-10   .35604-05
      CAHCO3+                   .00000     .00000     .00000     .20252-04   .11502-05   .12781-03
      CASO4(L)                  .00000     .00000     .00000     .12848-03   .84880-02   .70496-02
      CANO3+                    .00000     .00000     .00000     .63703-06   .34739-04   .24869-04
      MG++                      .00000     .00000     .00000     .41716-03   .39944-02   .35639-02
      MGOH+                     .00000     .00000     .00000     .22014-06   .46847-09   .37181-06
      MGSO3(L)                  .00000     .00000     .00000     .75013-06   .64736-05   .72042-04
      MGHCO3+                   .00000     .00000     .00000     .41764-05   .13479-06   .20058-04
      MGSO4(L)                  .00000     .00000     .00000     .66002-04   .24778-02   .27559-02
      MGCO3(L)                  .00000     .00000     .00000     .17243-05   .14335-10   .18725-05
      NA+                       .00000     .00000     .00000     .15771-02   .90278-02   .88922-02
      NAOH(L)                   .00000     .00000     .00000     .48581-09   .72494-12   .62610-09
      NACO3-                    .00000     .00000     .00000     .25701-06   .21385-11   .29008-06
      NAHCO3(L)                 .00000     .00000     .00000     .17126-05   .38759-07   .62764-05
      NASO4-                    .00000     .00000     .00000     .61576-05   .23137-03   .26723-03
      NANO3(L)                  .00000     .00000     .00000     .10454-06   .22718-05   .23700-05
      CL-                       .00000     .00000     .00000     .10309-02   .11001-01   .10881-01
    IONIC STRENGTH              .00000     .00000     .00000     .67685-02   .92283-01   .71357-01
    DISSOLVED SOLIDS (WPPM)     .00000     .00000     .00000     342.54      5799.3      4104.9
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07 AUG 75    21:25:32.870                       LG+E SIMULATION NO. 5
 STREAM NUMBER                     1            2            3            4            5            6
 SLURRY STREAM
   FLOW RATE  (LB/HR)           .00000       .00000       .00000       .00000      .65513+07    .66053+07
              (GAL/MIN)         .00000       .00000       .00000       .00000      12383.       12506.
   DENSITY    (LBS/GAL)         .00000       .00000       .00000       .00000      8.8179       8.8028
 SOLID PHASE
   FLOW RATE  (LBS/HR)          4212.1       720.00       59.007       .00000      .65356+06    .65833+06
              (LB-MOLES/HR)     56.855       9.4987       .77846       .00000      5285.7       5319.4
   ENTHALPY   (BTU/HR)        -.24051+08   -.41068+07   -.33934+06    .00000     -.29788+10   -.29988+10
   SPECIFIC GRAVITY             2.2401       3.2800       3.2800       .00000      2.5465       2.5453
   KEY COMP CONC (MOLE %)
     SO2                        .00000       .00000       .00000       .00000      36.244       36.291
     CO2                        .00000       .00000       .00000       .00000      .43476       .43095
     SO3                        .00000       .00000       .00000       .00000      2.1301       2.1024
     N2O5                       .00000       .00000       .00000       .00000      .00000       .00000
     CAO                        49.962       .00000       .00000       .00000      38.809       38.825
     MGO                        .38099-01    .00000       .00000       .00000      .00000       .00000
     NA2O                       .00000       .00000       .00000       .00000      .00000       .00000
     HCL                        .00000       .00000       .00000       .00000      .00000       .00000
     H2O                        50.000       .00000       .00000       .00000      22.382       22.350
   COMPOSITION(WT%)
     CAO                        .00000       .00000       .00000       .00000      .00000       .00000
     CA(OH)2                    99.940       .00000       .00000       .00000      .00000       .00000
     CACO3                      .00000       .00000       .00000       .00000      .78333       .77770
     CASO3                      .00000       .00000       .00000       .00000      .00000       .00000
     CASO3*1/2H2O               .00000       .00000       .00000       .00000      84.266       84.508
     CASO4                      .00000       .00000       .00000       .00000      .00000       .00000
     CASO4*2H2O                 .00000       .00000       .00000       .00000      6.6020       6.5264
     MGO                        .00000       .00000       .00000       .00000      .00000       .00000
     MG(OH)2                    .60000-01    .00000       .00000       .00000      .00000       .00000
     MGCO3                      .00000       .00000       .00000       .00000      .00000       .00000
     MGCO3*3H2O                 .00000       .00000       .00000       .00000      .00000       .00000
     MGCO3*5H2O                 .00000       .00000       .00000       .00000      .00000       .00000
     MGSO3                      .00000       .00000       .00000       .00000      .00000       .00000
     MGSO3*3H2O                 .00000       .00000       .00000       .00000      .00000       .00000
     MGSO3*6H2O                 .00000       .00000       .00000       .00000      .00000       .00000
     MGSO4                      .00000       .00000       .00000       .00000      .00000       .00000
     MGCL2                      .00000       .00000       .00000       .00000      .00000       .00000
     NA2O                       .00000       .00000       .00000       .00000      .00000       .00000
     NACL                       .00000       .00000       .00000       .00000      .00000       .00000
     INERTS                     .00000       100.00       100.00       .00000      8.3492       8.1882
     LIMESTONE                  .00000       .00000       .00000       .00000      .00000       .00000
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07 AUG 75    21:25:33.210                         LG+E SIMULATION NO.5
 STREAM NUMBER                        7           8           9           10          11
 TOTAL STREAM
    FLOW RATE (G/SEC)             10211.      7653.9      .83363+06   .84246+06   2551.3
              (G-MOLE/SEC)        516.99      423.68      42182.      42656.      92.976
    TEMPERATURE (DEG, K)          325.28      325.28      325.28      325.28      325.28
    PRESSURE  (ATM)               .00000      .00000      .00000      .00000      .00000
    ENTHALPY  (CAL/SEC)         -.37135+08  -.28777+08  -.30307+10  -.30639+10  -.83385+07
    SOLIDS    (WT %)              9.9667      .00000      9.9131      9.9667      40.000
              (MG/CU M)           .00000      .00000      .00000      .00000      .00000
 GAS PHASE
    FLOW RATE (G/SEC)             .00000      .00000      .00000      .00000      .00000
              (G-MOLE/SEC)        .00000      .00000      .00000      .00000      .00000
              (L/SEC AT T)        .00000      .00000      .00000      .00000      .00000
              (L/SEC, STP)        .00000      .00000      .00000      .00000      .00000
    ENTHALPY  (CAL/SEC)           .00000      .00000      .00000      .00000      .00000
    DENSITY   (G/L AT T)          .00000      .00000      .00000      .00000      .00000
    COMP      (G-MOLES/SEC)
      SO2                         .00000      .00000      .00000      .00000      .00000
      CO2                         .00000      .00000      .00000      .00000      .00000
      NO                          .00000      .00000      .00000      .00000      .00000
      NO2                         .00000      .00000      .00000      .00000      .00000
      O2                          .00000      .00000      .00000      .00000      .00000
      SO3                         .00000      .00000      .00000      .00000      .00000
      N2                          .00000      .00000      .00000      .00000      .00000
      HCL                         .00000      .00000      .00000      .00000      .00000
      H2O                         .00000      .00000      .00000      .00000      .00000
      CH4                         .00000      .00000      .00000      .00000      .00000
    COMP      (MOLE %)
      SO2                         .00000      .00000      .00000      .00000      .00000
      CO2                         .00000      .00000      .00000      .00000      .00000
      NO                          .00000      .00000      .00000      .00000      .00000
      NO2                         .00000      .00000      .00000      .00000      .00000
      O2                          .00000      .00000      .00000      .00000      .00000
      SO3                         .00000      .00000      .00000      .00000      .00000
      N2                          .00000      .00000      .00000      .00000      .00000
      HCL                         .00000      .00000      .00000      .00000      .00000
      H2O                         .00000      .00000      .00000      .00000      .00000
      CH4                         .00000      .00000      .00000      .00000      .00000
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07 AUG 75    21:25:33.412                          LG+E SIMULATION NO. 5
 STREAM NUMBER                         7           8            9            10           11
 SLURRY STREAM
   FLOW RATE (G/SEC)             10211.       .00000       .83363+06    .84246+06    2551.3
             (G-MOLE/SEC)        516.99       .00000       42182.       42656.       92.976
             (L/SEC AT T)        9.2753       .00000       756.43       765.26       1.5453
   ENTHALPY  (CAL/SEC)          -.37135+08    .00000      -.30307+10   -.30639+10   -.83385+07
   BULK DENSITY(G/ML AT T)       1100.8       .00000       1102.1       1100.9       1651.0
   KEY COMP RATE(GMOLES/SEC)
     SO2                         6.6687       .00000       550.23       550.23       6.6617
     CO2                         .94020-01    .00000       7.7445       7.7575       .80731-01
     SO3                         .60379       .00000       49.808       49.818       .43515
     N2O5                        .53435-02    .00000       .44001       .44088       .89090-03
     CAO                         7.3110       .00000       603.21       603.22       7.1684
     MGO                         .58722-01    .00000       4.8407       4.8451       .97843-02
     NA2O                        .41969-01    .00000       3.4560       3.4628       .69969-02
     HCL                         .99622-01    .00000       8.2106       8.2197       .16595-01
     H2O                         512.23       .00000       41773.       42263.       88.771
 LIQUID PHASE
   FLOW RATE (G/SEC)             9192.9       7653.9       .75099+06    .75850+06    1530.8
             (G-MOLES/SEC)       508.76       423.68       41513.       41978.       84.737
   ENTHALPY  (CAL/SEC)          -.34559+08   -.28777+08   -.28214+10   -.28514+10   -.57553+07
   DENSITY   (G/ML AT T)         .99115       .99086       .99281       .99115       .99086
   KEY COMP RATE(G-MOLES/SEC)
     SO2                         .10033-01    .35686-02    11.368       .82783       .71372-03
     CO2                         .14949-01    .13241-01    .45055       1.2334       .26483-02
     SO3                         .21805       .16839       18.229       17.991       .33677-01
     N2O5                        .53435-02    .44545-02    .44001       .44088       .89090-03
     CAO                         .18751       .13892       25.479       15.471       .27783-01
     MGO                         .58722-01    .48922-01    4.8407       4.8451       .97843-02
     NA2O                        .41969-01    .34985-01    3.4560       3.4628       .69969-02
     HCL                         .99622-01    .82975-01    8.2106       8.2197       .16595-01
     H2O                         508.13       423.19       41441.       41925.       84.638
   KEY COMP CONC(G-MOLES/KG H2O)
     SO2                         .10959-02    .46803-03    .15225-01    .10959-02    .46803-03
     CO2                         .16329-02    .17366-02    .60343-03    .16329-02    .17366-02
     SO3                         .23817-01    .22084-01    .24414-01    .23817-01    .22084-01
     N2O5                        .58366-03    .58421-03    .58931-03    .58366-03    .58421-03
     CAO                         .20481-01    .18219-01    .34123-01    .20481-01    .18219-01
     MGO                         .64141-02    .64161-02    .64831-02    .64141-02    .64161-02
     NA2O                        .45842-02    .45883-02    .46286-02    .45842-02    .45883-02
     HCL                         .10881-01    .10882-01    .10996-01    .10881-01    .10882-01
   PH                            6.7260       6.7267       7.0595       6.7260       6.7267
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07 AUG 75    21:25:33.652                      LG+E SIMULATION NO.5
 STREAM NUMBER                  7          8          9          10          11
 LIQUID PHASE(CONT)
   LIQ. H2O RATE (KG/SEC)    9.1552     7.6248     746.66     755.38     1.5250
              (LITERS/SEC)   9.2369     7.6951     752.06     762.12     1.5390
   CONC          (G-MOLES/KG H2O)
     H+                      .22511-06  .22428-06  .10479-06  .22511-06  .22428-06
     OH-                     .41778-06  .41665-06  .90712-06  .41778-06  .41665-06
     HSO3-                   .23004-03  .10216-03  .15317-02  .23004-03  .10216-03
     SO3=                    .10299-03  .45155-04  .15163-02  .10299-03  .45155-04
     SO4=                    .13744-01  .12833-01  .13932-01  .13744-01  .12833-01
     HCO3-                   .11459-02  .12240-02  .46686-03  .11459-02  .12240-02
     CO3=                    .85189-06  .89835-06  .76742-06  .85189-06  .89835-06
     NO3-                    .11401-02  .11427-02  .11477-02  .11401-02  .11427-02
     HSO4-                   .25353-06  .24089-06  .11523-06  .25353-06  .24089-06
     H2SO3(L)                .44774-08  .19961-08  .13697-07  .44774-08  .19961-08
     H2CO3(L)                .32623-03  .34982-03  .61063-04  .32623-03  .34982-03
     CA++                    .12584-01  .11505-01  .14898-01  .12584-01  .11505-01
     CAOH+                   .59666-07  .55351-07  .14878-06  .59666-07  .55351-07
     CASO3(L)                .69082-03  .28743-03  .11274-01  .69082-03  .28743-03
     CACO3(L)                .35604-05  .35629-05  .35552-05  .35604-05  .35629-05
     CAHCO3+                 .12781-03  .12705-03  .59749-04  .12781-03  .12705-03
     CASO4(L)                .70496-02  .62727-02  .78597-02  .70496-02  .62727-02
     CANO3+                  .24869-04  .23266-04  .28569-04  .24869-04  .23266-04
     MG++                    .35639-02  .36354-02  .32117-02  .35639-02  .36354-02
     MGOH+                   .37181-06  .38299-06  .71191-06  .37181-06  .38299-06
     MGSO3(L)                .72042-04  .33283-04  .90275-03  .72042-04  .33283-04
     MGHCO3+                 .20058-04  .22139-04  .71999-05  .20058-04  .22139-04
     MGSO4(L)                .27559-02  .27228-02  .23592-02  .27559-02  .27228-02
     MGCO3(L)                .18725-05  .20806-05  .14357-05  .18725-05  .20806-05
     NA+                     .88922-02  .89112-02  .89892-02  .88922-02  .89112-02
     NAOH(L)                 .62610-09  .63148-09  .13524-08  .62610-09  .63148-09
     NACO3-                  .29008-06  .31240-06  .25546-06  .29008-06  .31240-06
     NAHCO3(L)               .62764-05  .67828-05  .25416-05  .62764-05  .67828-05
     NASO4-                  .26723-03  .25590-03  .26277-03  .26723-03  .25590-03
     NANO3(L)                .23700-05  .24106-05  .23584-05  .23700-05  .24106-05
     CL-                     .10881-01  .10882-01  .10996-01  .10881-01  .10882-01
   IONIC STRENGTH            .71357-01  .67384-01  .78863-01  .71357-01  .67384-01
   DISSOLVED SOLIDS (WPPM)   4104.9     3806.5     5770.5     4104.9     3806.5
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07 AUG 75    21:25:33.895                          LG+E SIMULATION NO.5
 STREAM NUMBER                       7           8            9            10            11
 SOLID PHASE
   FLOW RATE (G/SEC)              1017.7      .00000       82639.       83966.        1020.5
             (G-MOLE/SEC)         8.2228      .00000       668.44       678.45        8.2393
   ENTHALPY  (CAL/SEC)         -.25753+07     .00000     -.20926+09   -.21249+09    -.25832+07
   SPECIFIC GRAVITY               2.5453      .00000       2.5465       2.5453        2.5446
   KEY COMP RATE(GMOLES/SEC)
     SO2                          6.6587      .00000       538.86       549.40        6.6610
     CO2                         .79071-01    .00000       7.2939       6.5240       .78083-01
     SO3                          .38574      .00000       31.579       31.827        .40147
     N2O5                         .00000      .00000       .00000       .00000        .00000
     CAO                          7.1235      .00000       577.73       587.75        7.1406
     MGO                          .00000      .00000       .00000       .00000        .00000
     NA2O                         .00000      .00000       .00000       .00000        .00000
     HCL                          .00000      .00000       .00000       .00000        .00000
     H2O                          4.1008      .00000       332.59       338.35        4.1335
   COMP RATE     (G MOLES/SEC)
     CAO                          .00000      .00000       .00000       .00000        .00000
     CA(OH)2                      .00000      .00000       .00000       .00000        .00000
     CACO3                       .79071-01    .00000       7.2939       6.5240       .78083-01
     CASO3                        .00000      .00000       .00000       .00000        .00000
     CASO3*1/2H2O                 6.6587      .00000       538.86       549.40        6.6610
     CASO4                        .00000      .00000       .00000       .00000        .00000
     CASO4*2H2O                   .38574      .00000       31.579       31.827        .40147
     MGO                          .00000      .00000       .00000       .00000        .00000
     MG(OH)2                      .00000      .00000       .00000       .00000        .00000
     MGCO3                        .00000      .00000       .00000       .00000        .00000
     MGCO3*3H2O                   .00000      .00000       .00000       .00000        .00000
     MGCO3*5H2O                   .00000      .00000       .00000       .00000        .00000
     MGSO3                        .00000      .00000       .00000       .00000        .00000
     MGSO3*3H2O                   .00000      .00000       .00000       .00000        .00000
     MGSO3*6H2O                   .00000      .00000       .00000       .00000        .00000
     MGSO4                        .00000      .00000       .00000       .00000        .00000
     MGCL2                        .00000      .00000       .00000       .00000        .00000
     NA2O                         .00000      .00000       .00000       .00000        .00000
     NACL                         .00000      .00000       .00000       .00000        .00000
     INERTS                       1.0993      .00000       90.702       90.702        1.0987
     LIMESTONE                    .00000      .00000       .00000       .00000        .00000
   XLS                            .00000      .00000       .00000       .00000        .00000
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07 AUG 75    21:25:34.108                              LG+E SIMULATION NO.5
 STREAM NUMBER                     7           8           9          10          11
 ELEMENTS(G-ATOMS/SEC
 TOTAL STREAM
      H                       1024.6      846.46      83555.      84535.      177.56
      C                       .94020-01   .13241-01   7.7445      7.7575      .80731-01
      N                       .10687-01   .89089-02   .88003      .88177      .17818-02
      O                       535.00      423.97      43653.      44143.      110.75
      NA                      .83937-01   .69969-01   6.9119      6.9256      .13994-01
      MG                      .58722-01   .48922-01   4.8407      4.8451      .97843-02
      S                       7.2725      .17196      600.04      600.05      7.0969
      CL                      .99622-01   .82975-01   8.2106      8.2197      .16595-01
      CA                      7.3110      .13892      603.21      603.22      7.1684
 GAS PHASE
      H                       .00000      .00000      .00000      .00000      .00000
      C                       .00000      .00000      .00000      .00000      .00000
      N                       .00000      .00000      .00000      .00000      .00000
      O                       .00000      .00000      .00000      .00000      .00000
      S                       .00000      .00000      .00000      .00000      .00000
      CL                      .00000      .00000      .00000      .00000      .00000
 LIQUID PHASE
      H                       1016.4      846.46      82890.      83858.      169.29
      C                       .14949-01   .13241-01   .45055      1.2334      .26483-02
      N                       .10687-01   .89089-02   .88003      .88177      .17818-02
      O                       509.15      423.97      41555.      42009.      84.795
      NA                      .83937-01   .69969-01   6.9119      6.9256      .13994-01
      MG                      .58722-01   .48922-01   4.8407      4.8451      .97843-02
      S                       .22808      .17196      29.596      18.819      .34391-01
      CL                      .99622-01   .82975-01   8.2106      8.2197      .16595-01
      CA                      .18751      .13892      25.479      15.471      .27783-01
 SOLID PHASE
      H                       8.2017      .00000      665.18      676.71      8.2669
      C                       .79071-01   .00000      7.2939      6.5240      .78083-01
      N                       .00000      .00000      .00000      .00000      .00000
      O                       25.857      .00000      2097.4      2133.4      25.957
      NA                      .00000      .00000      .00000      .00000      .00000
      MG                      .00000      .00000      .00000      .00000      .00000
      S                       7.0444      .00000      570.44      581.23      7.0625
      CL                      .00000      .00000      .00000      .00000      .00000
      CA                      7.1235      .00000      577.73      587.75      7.1406
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07 AUG 75    21:25:34.331                        LG+E SIMULATION NO.5
 STREAM NUMBER                        7           8            9           10           11
 TOTAL STREAM
    FLOW RATE (LBS/HR)            81037.      60746.       .66162+07   .66863+07    20249.
              (LB-MOLES/HR)       4103.1      3362.6       .33478+06   .33855+06    737.92
    TEMPERATURE(DEG F)            125.82      125.82       125.82      125.82       125.82
    PRESSURE  (PSIA)              .00000      .00000       .00000      .00000       .00000
    ENTHALPY  (BTU/HR)          -.53050+09  -.41110+09   -.43296+11  -.43771+11   -.11912+09
    SOLIDS    (WT % )             9.9667      .00000       9.9131      9.9667       40.000
              (GR/ACF, WET)       .00000      .00000       .00000      .00000       .00000
 GAS PHASE
    FLOW RATE (LBS/HR)            .00000      .00000       .00000      .00000       .00000
              (LB-MOLES/HR)       .00000      .00000       .00000      .00000       .00000
              (ACFM)              .00000      .00000       .00000      .00000       .00000
              (SCFM)              .00000      .00000       .00000      .00000       .00000
    ENTHALPY  (BTU/HR)            .00000      .00000       .00000      .00000       .00000
    DENSITY   (LBS/FT3)           .00000      .00000       .00000      .00000       .00000
    COMP      (MOLE %)
      SO2                         .00000      .00000       .00000      .00000       .00000
      CO2                         .00000      .00000       .00000      .00000       .00000
      NO                          .00000      .00000       .00000      .00000       .00000
      NO2                         .00000      .00000       .00000      .00000       .00000
      O2                          .00000      .00000       .00000      .00000       .00000
      SO3                         .00000      .00000       .00000      .00000       .00000
      N2                          .00000      .00000       .00000      .00000       .00000
      HCL                         .00000      .00000       .00000      .00000       .00000
      H2O                         .00000      .00000       .00000      .00000       .00000
      CH4                         .00000      .00000       .00000      .00000       .00000
 LIQUID PHASE
    FLOW RATE (LBS/HR)            72961.      60746.       .59604+07   .60199+07    12149.
              (LB-MOLES/HR)       4037.9      3362.6       .32948+06   .33316+06    672.53
    ENTHALPY  (BTU/HR)          -.49371+09  -.41110+09   -.40307+11  -.40736+11   -.82220+08
    DENSITY   (LBS/GAL)           8.2712      8.2687       8.2850      8.2712       8.2687
    KEY COMP CONC(MOLALITY)
      SO2                         .10959-02   .46803-03    .15225-01   .10959-02    .46803-03
      CO2                         .16329-02   .17366-02    .60343-03   .16329-02    .17366-02
      SO3                         .23817-01   .22084-01    .24414-01   .23817-01    .22084-01
      N2O5                        .58366-03   .58421-03    .58931-03   .58366-03    .58421-03
      CAO                         .20481-01   .18219-01    .34123-01   .20481-01    .18219-01
      MGO                         .64141-02   .64161-02    .64831-02   .64141-02    .64161-02
      NA2O                        .45842-02   .45883-02    .46286-02   .45842-02    .45883-02
      HCL                         .10881-01   .10882-01    .10996-01   .10881-01    .10882-01
    PH                            6.7260      6.7267       7.0595      6.7260       6.7267
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07 AUG 75    21:25:34.557                           LG+E SIMULATION NO.5
 STREAM NUMBER                  7            8            9            10           11
 LIQUID PHASE(CONT)
    LIQ H2O RATE ( LBS/HR)   72661.       60515.       .59260+07    .59952+07    12103.
                 (GAL/MIN)   147.02       122.44       11990.       12130.       24.488
    CONC (MOLALITY)
      H+                     .22511-06    .22428-06    .10479-06    .22511-06    .22428-06
      OH-                    .41778-06    .41665-06    .90712-06    .41778-06    .41665-06
      HSO3-                  .23004-03    .10216-03    .15317-02    .23004-03    .10216-03
      SO3=                   .10299-03    .45155-04    .15163-02    .10299-03    .45155-04
      SO4=                   .13744-01    .12833-01    .13932-01    .13744-01    .12833-01
      HCO3-                  .11459-02    .12240-02    .46686-03    .11459-02    .12240-02
      CO3=                   .85189-06    .89835-06    .76742-06    .85189-06    .89835-06
      NO3-                   .11401-02    .11427-02    .11477-02    .11401-02    .11427-02
      HSO4-                  .25353-06    .24089-06    .11523-06    .25353-06    .24089-06
      H2SO3(L)               .44774-08    .19961-08    .13697-07    .44774-08    .19961-08
      H2CO3(L)               .32623-03    .34982-03    .61063-04    .32623-03    .34982-03
      CA++                   .12584-01    .11505-01    .14898-01    .12584-01    .11505-01
      CAOH+                  .59666-07    .55351-07    .14878-06    .59666-07    .55351-07
      CASO3(L)               .69082-03    .28743-03    .11274-01    .69082-03    .28743-03
      CACO3(L)               .35604-05    .35629-05    .35552-05    .35604-05    .35629-05
      CAHCO3+                .12781-03    .12705-03    .59749-04    .12781-03    .12705-03
      CASO4(L)               .70496-02    .62727-02    .78597-02    .70496-02    .62727-02
      CANO3+                 .24869-04    .23266-04    .28569-04    .24869-04    .23266-04
      MG++                   .35639-02    .36354-02    .32117-02    .35639-02    .36354-02
      MGOH+                  .37181-06    .38299-06    .71191-06    .37181-06    .38299-06
      MGSO3(L)               .72042-04    .33283-04    .90275-03    .72042-04    .33283-04
      MGHCO3+                .20058-04    .22139-04    .71999-05    .20058-04    .22139-04
      MGSO4(L)               .27559-02    .27228-02    .23592-02    .27559-02    .27228-02
      MGCO3(L)               .18725-05    .20806-05    .14357-05    .18725-05    .20806-05
      NA+                    .88922-02    .89112-02    .89892-02    .88922-02    .89112-02
      NAOH(L)                .62610-09    .63148-09    .13524-08    .62610-09    .63148-09
      NACO3-                 .29008-06    .31240-06    .25546-06    .29008-06    .31240-06
      NAHCO3(L)              .62764-05    .67828-05    .25416-05    .62764-05    .67828-05
      NASO4-                 .26723-03    .25590-03    .26277-03    .26723-03    .25590-03
      NANO3(L)               .23700-05    .24106-05    .23584-05    .23700-05    .24106-05
      CL-                    .10881-01    .10882-01    .10996-01    .10881-01    .10882-01
    IONIC STRENGTH           .71357-01    .67384-01    .78863-01    .71357-01    .67384-01
    DISSOLVED SOLIDS (WPPM)  4104.9       3806.5       5770.5       4104.9       3806.5
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07 AUG 75    21:25:34.794                          LG+E SIMULATION NO.5
 STREAM NUMBER                    7           8            9            10           11
 SLURRY STREAM
   FLOW RATE (LB/HR)            81037.       .00000       .66162+07    .66863+07    20249.
             (GAL/MIN)          153.43       .00000       12511.       12659.       30.927
   DENSITY   (LBS/GAL)          8.8028       .00000       8.8141       8.8028       10.912
 SOLID PHASE
   FLOW RATE (LBS/HR)           8076.8       .00000       .65587+06    .66640+06    8099.5
             (LB-MOLES/HR)      65.261       .00000       5305.1       5384.6       65.392
   ENTHALPY  (BTU/HR)         -.36791+08     .00000      -.29895+10   -.30356+10   -.36904+08
   SPECIFIC GRAVITY             2.5453       .00000       2.5465       2.5453       2.5446
   KEY COMP CONC (MOLE %)
     SO2                        36.291       .00000       36.212       36.291       36.172
     CO2                        .43095       .00000       .49016       .43095       .42403
     SO3                        2.1024       .00000       2.1222       2.1024       2.1802
     N2O5                       .00000       .00000       .00000       .00000       .00000
     CAO                        38.825       .00000       38.825       38.825       38.777
     MGO                        .00000       .00000       .00000       .00000       .00000
     NA2O                       .00000       .00000       .00000       .00000       .00000
     HCL                        .00000       .00000       .00000       .00000       .00000
     H2O                        22.350       .00000       22.351       22.350       22.447
   COMPOSITION(WT%)
     CAO                        .00000       .00000       .00000       .00000       .00000
     CA(OH)2                    .00000       .00000       .00000       .00000       .00000
     CACO3                      .77770       .00000       .88343       .77770       .76583
     CASO3                      .00000       .00000       .00000       .00000       .00000
     CASO3*1/2H2O               84.508       .00000       84.217       84.508       84.300
     CASO4                      .00000       .00000       .00000       .00000       .00000
     CASO4*2H2O                 6.5264       .00000       6.5796       6.5264       6.7735
     MGO                        .00000       .00000       .00000       .00000       .00000
     MG(OH)2                    .00000       .00000       .00000       .00000       .00000
     MGCO3                      .00000       .00000       .00000       .00000       .00000
     MGCO3*3H2O                 .00000       .00000       .00000       .00000       .00000
     MGCO3*5H2O                 .00000       .00000       .00000       .00000       .00000
     MGSO3                      .00000       .00000       .00000       .00000       .00000
     MGSO3*3H2O                 .00000       .00000       .00000       .00000       .00000
     MGSO3*6H2O                 .00000       .00000       .00000       .00000       .00000
     MGSO4                      .00000       .00000       .00000       .00000       .00000
     MGCL2                      .00000       .00000       .00000       .00000       .00000
     NA2O                       .00000       .00000       .00000       .00000       .00000
     NACL                       .00000       .00000       .00000       .00000       .00000
     INERTS                     8.1882       .00000       8.3197       8.1882       8.1609
     LIMESTONE                  .00000       .00000       .00000       .00000       .00000
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